THERMAL POWER GENERATION WITHOUT CO_2 EMISSIONS. AN ASSESSMENT OF ADDITIONAL COSTS IN ARGENTINA

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INTRODUCTION

A study was undertaken to explore two options of CO_2 capture from natural gas combined cycles (NGCC) in Argentina. Natural gas is the predominant fuel for thermal power generation in the country and NGCC has been the choice for new generating units, particularly after the liberalization of the electricity sector in 1992. A high level assessment on sources to sinks matching was done taking into account the technology type, location, emissions and vintage of large stationary CO_2 sources and the characteristics of the local sedimentary basins. Based on the proximity of thermal power plants to geologic reservoirs with potential for sequestering large volumes of CO_2, a NGCC was selected as reference plant and a gas field for geological storage site. The case study concerns the conceptual design of post and pre-combustion capture systems; compression-dehydration units for each of the CO_2 rich stream leaving the capture systems, pipeline transportation and injection.

The same hourly flow rate of natural gas was employed to design the associated post and pre-combustion facilities using today's best available technology. A monoethanolamine process was selected for post-combustion capture and a steam methane reformer coupled with a water-gas shift reactor for hydrogen generation in the pre-combustion facility. An extensive survey of published data and models concerning physical equilibria, kinetics and heat transfer of the processes involved was conducted. Commercial process simulators were used in the design of the capture systems and the CO_2 supercritical pipeline as well as the compression-dehydration unit. The results were as far as possible validated in regard to the reviewed literature.

To the best of our knowledge this is the first assessment undertaken in Argentina on CO_2 capture and storage from thermal power plants. We thus have assessed the main technological aspects of the CO_2 recovery units to derive not only the building blocks for local cost estimates but also to provide detailed information on process conditions, energy requirements and equipment size. For countries involved for the first time in this type of assessment it is crucial to develop local modelling capabilities because the design of post and pre-combustion units applied to NGCC was found to depend strongly on the assumptions about the reference plant and the details of the CO_2 capture and storage systems.

The Argentinean energy system

In 2000, gas accounted for 44.7% of Argentina's primary energy supply. Oil, the next largest source of energy, was responsible for nearly 38%. Hydroelectric power accounted for 5.3%, while nuclear power made up almost 2% of the country's total energy supply. The remaining primary energy was supplied from biomass (mainly wood and bagasse). For the purposes of CO_2 capture and storage, Fig. 1 shows the energy flows through the Argentinean economy in terms of carbon flows. It is widely accepted that in the near term, CO_2 will be captured from sources emitting large quantities at fixed locations. In Argentina, this type of recoverable CO_2 is most likely to come from natural gas burned for electricity or heat generation in the energy and manufacturing industries. In these sectors, a lesser amount of CO_2 arises from the burning of liquid fuels (~13%) and coal (~6%). About 1000 Gg of CO_2 (equivalent to 273 Gg of C, not shown in Fig. 1) are already recovered from natural gas sweetening in the oil and gas industry. This CO_2 that is presently vented to the atmosphere constitute an early opportunity if capture and storage becomes an attractive option for the country. On the other hand, the numerous CO_2 stationary sources in the commercial and

residential are generally much smaller and less attractive for CO_2 capture. Similarly, the mobile characteristics of the CO_2 sources of the transportation sector make them less amenable for recovering CO_2.

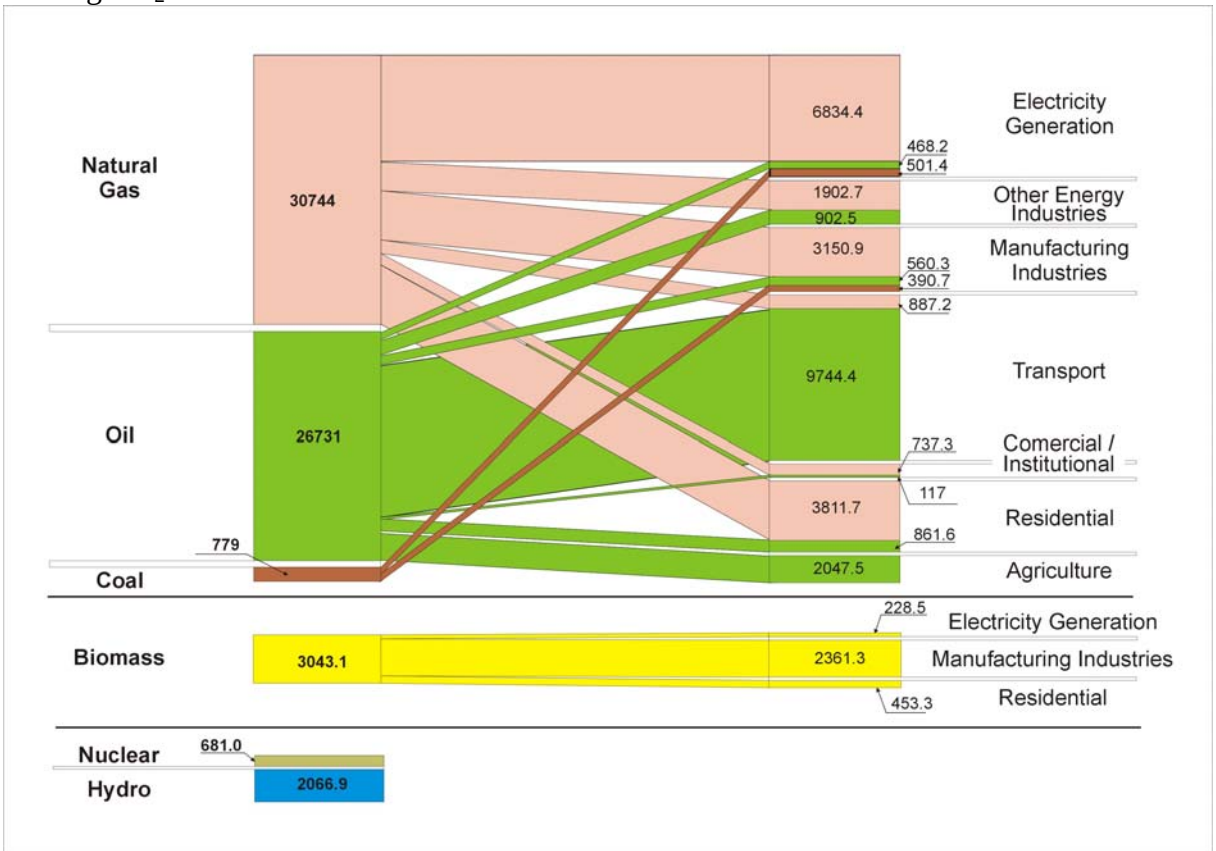

Fig. 1. Carbon flows (Gg of C) in the Argentinean energy system

Oil and gas system

There are 24 known sedimentary basins in the Argentinean territory. Eleven of these are located entirely onshore, six are combined onshore/offshore and seven are entirely offshore. Total onshore surface comprises 1.36 10^6 km^2, and total offshore surface includes 0.44 10^6 km^2 on the South Atlantic shelf. A substantial portion of the 1.80 10^6 km^2 in sedimentary basins has yet to be evaluated by exploratory drilling. Only five of these basins have seen commercial production of oil and gas. In the northern and central regions of the country, there are three basins located on the West: Northwest, Cuyana and Neuquén. The other two productive basins are located in Patagonia and include onshore and offshore fields: San Jorge and Austral. Proven reserves for year 2001 were about 460 10^6 m^3 of oil and 760 10^9 m^3 of natural gas. Neuquén, located in the central part of the country, is Argentina's most productive basin and has about 50% of the country's total remaining hydrocarbon reserves (both oil and gas). The country has an extensive oil and gas pipeline network with a radial structure centred in the metropolitan area of Buenos Aires, the region with the highest consumption. In addition, oil lines connect to export ports on the Atlantic and Pacific coasts and gas pipelines to Bolivia, Brazil, Chile and Uruguay.

The electricity system

Argentina had in 2000 an electricity consumption of about 71 TWh. About 51% of the generation capacity is based on thermal power plants, 44% belongs to hydropower facilities and 5% is nuclear power. The country has two interconnected systems that are independent. The main system, comprising the central and northern part of the country, is referred as the Wholesale Electricity Market (WEM) while the Wholesale Electricity Market of the Patagonian System (WEMPS) located in the southern part of the country is much smaller than the first one.

Eight regions are defined in the WEM according to supply and demand concentrations: Buenos Aires, Centre, Comahue, Cuyo, Great Buenos Aires (GBA), Litoral, North-eastern (NEA) and

North-western (NOA). The GBA region encompasses the metropolitan area of Buenos Aires and constitutes the region of highest demand. Power demand from this area is almost 48% of the total demand of the country while its installed capacity is entirely thermal and corresponds to 24% of the total. The Comahue region (composed by the provinces of La Pampa, Neuquén and Rio Negro) is the main supplier of electricity from both hydro and thermal power. There is a main high voltage transmission grid (in 500 kV) whose length is over 9,000 km operated by a company which interconnects several electricity regions. Each of these geographical regions is served by a regional transmission company in lower high voltages lines (from 330 kV to 132 kV).

Up to the 70's, the electricity generation sector of Argentina was largely based on conventional thermal power. A substitution policy of thermal power by hydro and nuclear power led to its diversification and a more balanced energy mix.**¡Error! No se encuentra el origen de la referencia.**[1] The role of the State in defining the energy supply structure, as well as the instruments for policies implementation has completely changed since the reorganisation and privatisation of the energy industries in 1992. Market reforms have included private actors in the electricity sector that have incorporated their priorities into key decisions that affect the environment. Private generators chose to expand the installed capacity based on natural gas combined cycles. The trend towards combined-cycle generation supply increased from less than 1% of the total installed capacity in 1992 up to 27% in 2003. This was possible through the addition of new plants or the integration of gas turbines to combined-cycle units.

Thermal power is generated primarily from natural gas; liquid fuels are used only in four regions of the WEM mostly during winter when residential consumption is given priority. Coal is burned in only one thermal power plant. Natural gas share has increased from 70% in 1992 up to about 98% in 2003. This fuel consumption pattern of the electricity sector of Argentina does not allow a wide margin for fuel switching if reductions in CO_2 emissions are sought. In this case, increasing the share of hydro, nuclear and renewable power as well as CO_2 capture and storage are the options left for the country.

CO_2 SOURCES AND POTENTIAL GEOLOGICAL STORAGE IN ARGENTINA

Almost 56 Tg of CO_2 were emitted in 2000 from large stationary sources in Argentina, namely those emitting over 100 Gg of CO_2 per year. The largest amounts of these emissions originated from: natural gas combustion in power plants (44%) > cement industry (16%) > iron and steel industry (14%) > fuel combustion in oil refineries (13%). Comparatively negligible quantities originated from: ammonia production (0.06%) < hydrogen production (0.1%) < natural gas processing (1.7%).

Emissions from power plants

The overall CO_2 emissions of the electricity sector amounted to 24.6 Tg in year 2000. The four regions of relatively high emissions are (in Tg CO_2): Great Buenos Aires (8.1) > Buenos Aires (4.8) > Comahue (4.0) > North-western (3.3). The CO_2 emissions in the North-eastern region are negligible as electricity is mostly generated from hydro power there. About half of the emissions arising from natural gas belongs to gas turbines and NGCC and has an average CO_2 content between 3 and 4% (dry basis). The other half belongs to boilers with a CO_2 content between 6 and 9%. Emissions from liquid or solid fuels have a higher CO_2 content in the range 14 to 16%.

Although our work focuses on CO_2 emissions from thermal power plants, from a broader perspective it is worth considering other emission sources that are also candidate for capture and storage in Argentina. Natural gas burning in power plants accounts for the largest amount of CO_2 emitted and the largest number of sources (61) but its CO_2 content is the lowest (~3%). The cement industry with 14 large sources and a mean CO_2 content of 20% and the iron and steel industry with 5 large sources and a typical CO2 content of 15% may also be attractive candidates for CO_2 capture. Although the amounts of CO_2 that arise from ammonia, hydrogen and natural gas sweetening processes are comparatively low, these streams of almost pure CO_2 represent early opportunities for storage since the capture process is already done because of the purity requirements of the main products.

Potential geological storage sites

In this study, we have assumed that the CO_2 emitted by large sources belonging to the Argentinean electricity sector will be stored in geological reservoirs that are located relatively close (< 100 km) from the thermal power plant. Although the assessment of the suitability of the local geological media using a geosciences based analysis is beyond the scope of our work, we have attempted a screening of the Argentinean sedimentary basins. To this end, we have adopted the results from a recent study on geological storage prospectivity of CO_2 for the world's sedimentary basins.[2] Extrapolating the results of detailed assessments on the sedimentary basins of Australia[3] and Canada, [4] Bradshaw and Dance[2] were able to categorize the potential for CO_2 storage of major and minor hydrocarbon provinces as well as non-petroliferous sedimentary basins using the 2000 worldwide assessment of the potential for hydrocarbon prospectivity undertaken by the US Geological Survey (USGS).[5] At a very simplistic level, the basins were classified as *i*) highly prospective (those that are world class petroleum basins and are "high priority" or "frontier" basins for the USGS World Petroleum Assessment); *ii*) prospective (low to high, those that are minor petroleum basins but not world class, as well as other sedimentary basins that have not been highly deformed); *iii*) non prospective (those that are highly deformed sedimentary basins and other geological provinces principally comprising fold belts, metamorphic and igneous rocks). Large areas of three of the five productive sedimentary basins of Argentina, namely Austral, Neuquén and San Jorge were categorized as highly prospective for geological storage of CO_2 in the referenced study.[2] The North-western and Cuyana basins were classified as non-prospective while most of the non-productive sedimentary basins, except those located in the mountainous region of the Andes, were classified as prospective.

Sources and sinks matching

The geographical relationships between emission sources and potential storage sites were considered to assess the future implementation of the geological storage of CO_2. Information relevant to all the Argentinean thermal power plants were compiled and input to a dataset containing the location of the plant, type of fuel used, fuel consumption and associated CO_2 emissions for all years in the period 1999 – 2003. This information was used in a GIS application that allowed superimposing the loci of the thermal power plants on the map of both productive and non-productive sedimentary basins of Argentina (Fig. 2).

Large sources located within the productive basins that were ranked as highly prospective for the geological storage of CO_2 can be grouped in two clusters. One is composed by 5 thermal power plants (with individual emissions within the range 0.3 – 1.6 Tg CO_2/year), located in the Neuquén basin. The second cluster is formed by 2 thermal power plants in the San Jorge basin that have comparatively smaller individual emissions (below 0.3 Tg CO_2/year). Sources in the Austral basin emit annually less than 0.1 Tg CO_2 per plant and were consequently disregarded. One large source (1.8 Tg CO_2) is located in the Cuyana basin while 4 large sources (0.4 – 1.3 Tg CO_2/year) are within the North-western basin. However, these power plants were not taken into account in our study because of the classification as non-prospective assigned to the Cuyana and North-west basins. The largest sources pertaining to the thermal power plants of the metropolitan area of Buenos Aires (with individual emissions in the range 0.6 – 3 Tg CO_2/year) are located near a prospective non-productive sedimentary basin. In spite of their large emissions, these sources were also disregarded in our study since it is unlikely that a CO_2 capture and storage project would firstly occur near the densely populated metropolitan area of Buenos Aires.

Neuquén, located in the central part of the country and Argentina's most productive basin, was finally selected for our study. This basin, concentrates about 50% of the country's total remaining hydrocarbon reserves (both oil and gas), has an extensive oil and gas pipeline network and appropriate thickness for CO_2 storage (ranging between 500 m and 5000 m with median of 2500 m).[5] Some of the new thermal power units have been installed in this region after the liberalization of the electricity market. Private actors have found attractive to install generating machines close to the oil and gas fields to take advantage of lower fuel prices relative to negligible transport costs.

Furthermore, large stationary CO_2 sources belonging to natural gas sweetening and cement production are also present in this basin.

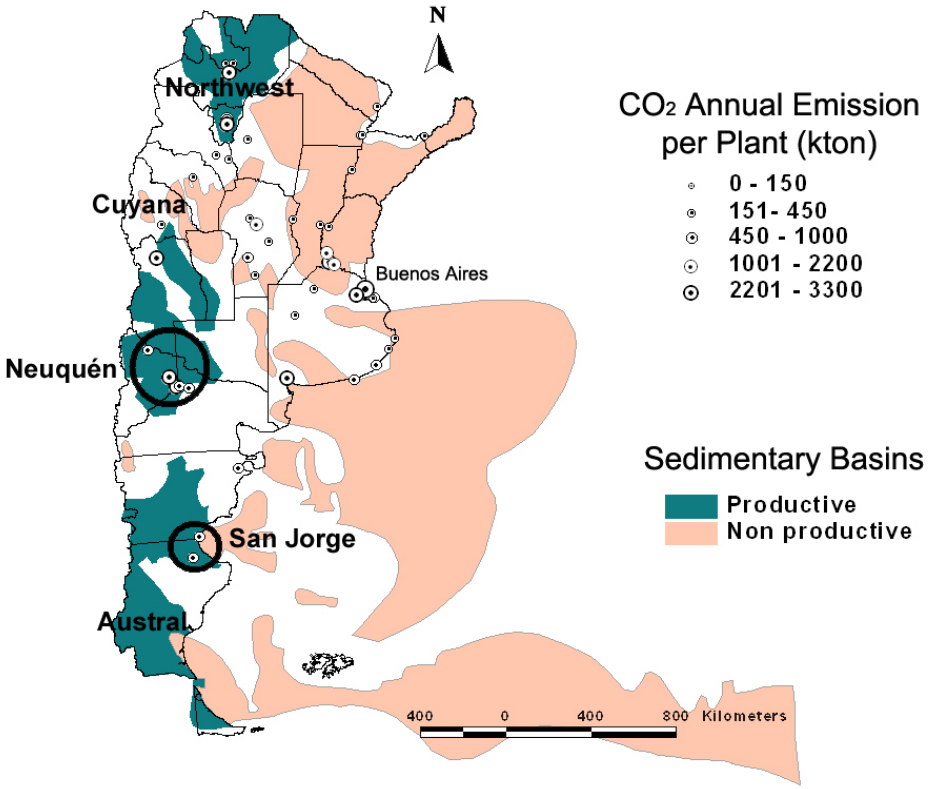

Fig. 2 Location of thermal power plants and sedimentary basins in Argentina

SELECTION OF A CASE STUDY

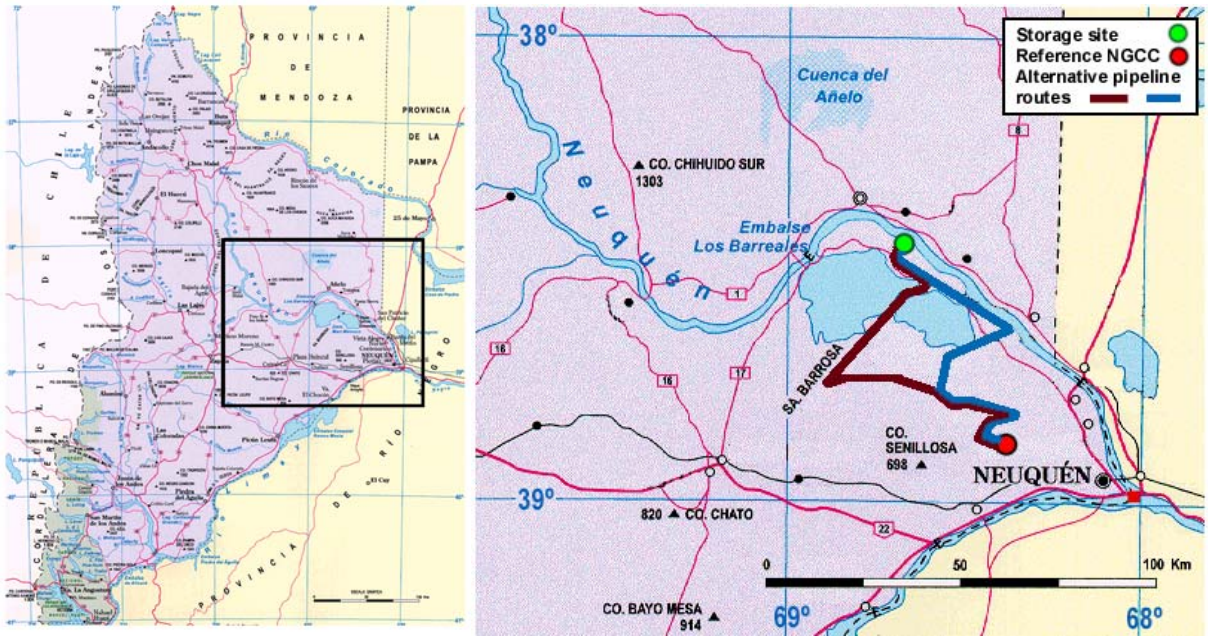

Fig. 3. Left: map showing Neuquén province where the reference power plant and storage site are located. Right: location of the facilities and the two alternative pipeline routes.

To focus on the main variables associate to CO_2 recovery and transmission and the Argentinean conditions, we chose to undertake a case study for a local thermal power plant. Based on the results of the exploration on source and sink matching, a NGCC installed in the Neuquén province was selected as reference plant for this case study. This reference NGCC is integrated by five gas turbines and one steam turbine. Its mean annual emission (2.1 Tg CO_2) represents more than 8% of the total emissions from the country's electricity sector. The Loma de la Lata gas field pertaining to

the Neuquén basin and located at ~ 60 km in a straight line from the power plant was selected as prospective storage site. The case study concerns the conceptual design of: *i*) post-combustion capture facilities, *ii*) pre-combustion capture facilities and *iii*) two alternative pipeline routes to transport the captured CO_2 to the storage site. Fig. 3 illustrates the location of the power plant, the storage site and the two alternative routes for the hypothetical pipeline.

Description of the power plant

The plant is located on the gas field that supplies the fuel used for electricity generation. It is composed by a 672 MW NGCC integrated by five 48 MW and one 131 MW gas turbines and one 301 MW vapour turbine. It was developed in four phases between 1993 and 1999 along which different generating machines were installed. Average effluent gas conditions are 96–110 °C; 13.5–14.2% of oxygen and about 4% of CO_2 (dry basis).

A natural gas flow rate of 136,000 $Nm^3\ h^{-1}$ (equivalent to 1350 MW_{th} on a LHV basis) was adopted as the design basis. This value corresponds to the amount of natural gas burned under average operating conditions of the reference NGCC. It was calculated taken into account 1) the monthly electricity generated and nominal fuel consumption for the entire NGCC during the period 2000-2002; and 2) disaggregated information by turbine for year 2002, namely monthly electricity generated, fuel consumption and water consumption for the NO_x control system.

Assessment of technological aspects

To explore the key parameters that affect the performance and cost of CO_2 capture, transmission and storage, we have assessed the main technological aspects of these technologies for the reference plant. A number of previous studies have reported cost and performance data for post-combustion options for thermal power plants using amine-based absorption.[6-13] Pre-combustion options for large thermal power plants have also received attention.[14-16] Finally some studies deal with the comparative assessment of different options that apart from post and pre-combustion capture include other options such as oxy-fuel combustion.[17-18] It has been recently pointed out that CO_2 transmission, the critical linking component between capture and storage has been often overlooked as compared to the amount of research devoted to the capture and storage components of the process.[19] However, there is an extensive experience in CO_2 transmission linked to CO_2 injection as an enhanced oil-recovery (EOR) technique. Studies by several researchers have reached the conclusion that high pressure supercritical pipelines are the most economical transportation mode for the large volumes of CO_2 needed for EOR.[20-23] Key design considerations for CO_2 pipeline transmission and the difference with those relative to conventional natural gas or liquid pipeline have been discussed in several publications.[24-26] In spite of this increasing body of knowledge, as pointed out by Rao and Rubin,[27] detailed models of the overall system of capture, transmission and storage are generally not available and also, cost data reported in different studies tend to be limited and often incomplete. We aimed at establishing a framework for quantifying the key variables involved in post and pre-combustion CO_2 capture from thermal power plants as well as in CO_2 transmission via pipeline.

The three main challenges face by CO_2 capture from thermal power plants and particularly those burning natural gas such as the reference NGCC of our study concern: (*i*) the large volumes that must be handled, (*ii*) the very low driving force for mass transfer for CO_2 removal in post-combustion capture and (*iii*) the strongly endothermic characteristics of the reaction system that converts the stable CH_4 molecule into the more reactive synthesis gas (CO and H_2) used in pre-combustion capture. These aspects imply large equipment size and large energy requirements for both types of capture. The removal of CO_2 (and other acid gases such as H_2S) from gas streams is a common practice in the natural gas and ammonia industries, oil refineries, and petrochemical plants. Several techniques are available for commercial plants capturing CO_2 from power plant flue gas in post-combustion systems. However, absorption with alkanolamines, especially monoethanolamine (MEA) that has been developed since the 1930s is still the major industrial technology.[28] For natural gas based pre-combustion systems, steam methane reforming (SMR) is currently the largest and generally the most economical way to make synthesis gas.[29-30] The key challenge of this process is heat transfer for the strongly endothermic reforming reaction.[29] In spite of the availability of these

technologies, the unprecedented scale of CO_2 capture from power plant operations imposes such a

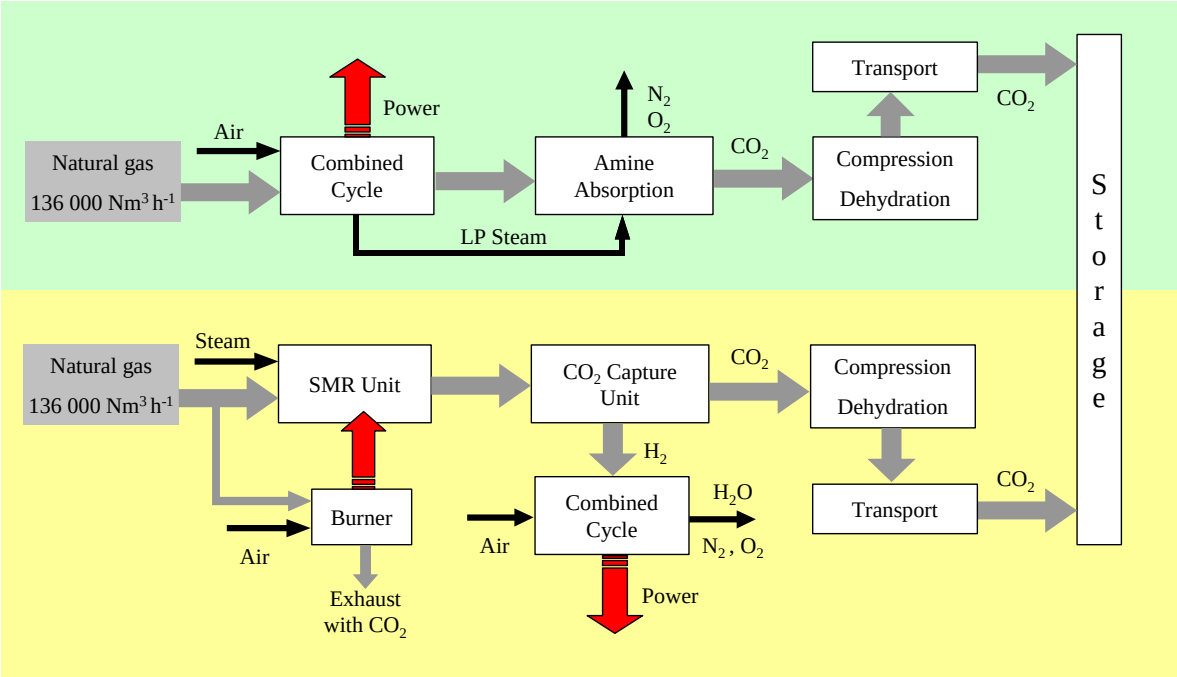

Fig. 4 Comparative assessment of capture and storage options

significant level of energy penalties and equipment size that in part explains the R&D efforts aimed at improving current industrial technology and the development of newer technology undertaken worldwide in the field of CO_2 capture.

The design framework of our study is depicted in Fig. 4. A MEA post-combustion capture system and a SMR pre-combustion capture system were designed for the same natural gas flow rate of 136,000 $Nm^3\ h^{-1}$. The amount of CO_2 captured in the post-combustion unit is larger than the amount captured in the pre-combustion system. In the last case, the exhaust gas from the combustion of natural gas in the furnace burner is not sent to the CO_2 capture unit because this gas is at a much lower partial pressure than that leaving the SMR unit. Compression-dehydration units and pipelines for the two alternative routes were designed for each CO_2 stream leaving post and pre-combustion units since these have different flow rates and compositions. The injection requirements for storage were also calculated.

POST-COMBUSTION CAPTURE FROM THE REFERENCE POWER PLANT

A Monoethanolamine recovery unit was designed to capture the CO_2 emitted by the reference NGCC. The properties of the flue gas from the NGCC that is input to the MEA unit are indicated in the process flow diagram of Fig. 5. Practically negligible amounts of NO_x and SO_2 are present in the flue gas. As design basis, we selected monthly average operating conditions (electricity generated: 411 GWh, power supplied: 606 MW), and calculated the flow rate (186,513 kmole/h) of the flue gas that is input to the recovery unit. The composition of the flue gas (in mole fraction) is: CO_2 (0.032), O_2 (0.137), N_2 (0.751) and H_2O (0.080).

Design of an amine-based process

The CO_2 removal efficiency is a function of parameters such as liquid to gas ratio (L/G), temperature, pressure, and MEA concentration of the liquid stream entering the absorber. The influence of these parameters was studied using a process simulator that was previously applied in assessing CO_2 removal by alkanolamines.[7, 31] Several steps were taken prior to the simulation runs, namely selecting the model for physical properties evaluation, verifying its results with published equilibrium data,[32-35] selecting test cases and verifying results. The main assumptions used were complete combustion, no pollutants in flue gases, adiabatic absorption and maximum MEA concentration in water (30% mass). Main variables for the absorber and the stripper are indicated in Fig. 5 The possibility of not feeding the entire flue gas stream leaving the NGCC was also assessed for CO_2 recoveries below 80%. A derivation ratio (ß) between the gas flow rate input to the MEA

unit and the flue gas rate was used as design variable. This option was analyzed to consider smaller equipment size for lower requirements. The influence of the key variables in the hydraulic design of the absorber and the stripper was also evaluated.

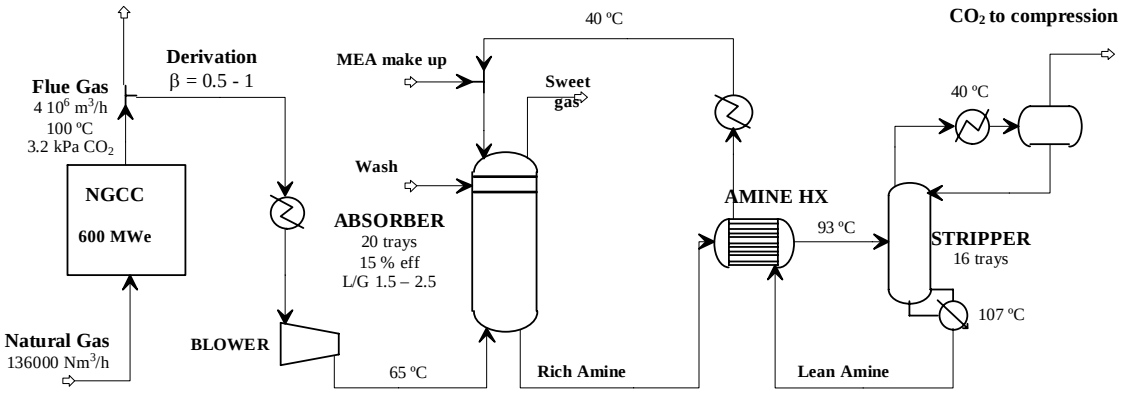

Fig. 5 Flow diagram of the post-combustion capture with main design parameters

The maximum CO_2 recovery achievable is 95%. The fraction of CO_2 recovered in the MEA unit is shown as a function of the stripper heat duty in Fig. 6a. The ratio (q) between this heat amount (MW_{th}) and the net power supplied by the NGCC (MW_e) was used as a normalized variable to show the significance of the energy required to regenerate the amine. For values of $q < 0.35$, the fraction of CO_2 recovered increases as the L/G ratio in the absorber increases. This situation reverts for higher heat duties. This behavior can be explained by the trade-off between the mass transfer driving force in the absorber and the reboiler heat duty that is linked to the re-circulation rate of the liquid streams in the unit. These opposing factors equilibrate for CO_2 recoveries of about 0.65 and $q \sim 0.35$ where the amount of CO_2 recovered is independent of the L/G

Besides the large heat duty required to regenerate the amine, another drawback of the process is the size of the absorption and stripping columns. For a maximum tray spacing of 36” (0.91 m), the required absorber diameter (D) is 23 m (or 6 columns, D=10 m) while for the stripping column D= 9-11 m, depending on the amount of heat used. Heights are of about 30 m for the absorber and 25 m for the stripper. Increasing the spacing between trays can reduce these values but this alternative requires experimental verification. Although the design was based on sieve trays, packed towers were also considered and no significant differences in tower diameter were encountered for random packings.

The analysis of the key variables of the MEA unit for values of the derivation ratio ß between 0.5 and 1 is shown in Fig. 6b for L/G ratios 1.5 and 2.25. The maximum achievable recovery is always 95% of the input CO_2. Similar amounts of CO_2 can be obtained for recoveries over 0.6 by adjusting the inlet gas flow rate and the L/G ratio. For CO_2 recoveries below 0.5 and fixed stripper heat duty, it is possible to increase the recovered amount of CO_2 by increasing the flow rate of the input gas stream and the L/G ratio. The absorber diameter as a function of ß with the tray spacing and number of columns as parameters was also explored. For a constant gas flow rate, the column diameter is inversely proportional to the square root of the number of columns. For a fixed number of columns, a linearly decreasing trend of the diameter with decreasing gas flow rates is observed. A comparative assessment shows that for high recoveries it is convenient to feed the entire flue gas stream to the MEA recovery unit since the reduction in column diameter is not significant. For validation purposes, our results were compared with previous work.[7-10] Our value of 4.5 MJ_{th}/kg of CO_2 for the ratio between the reboiler heat duty and the amount of CO_2 recovered for a recovery fraction of 0.85 is within the range 3-5 reported in the literature. The values of absorber diameters are in a satisfactory agreement with those obtained using the correlation suggested by Chapel *et al.*[8]

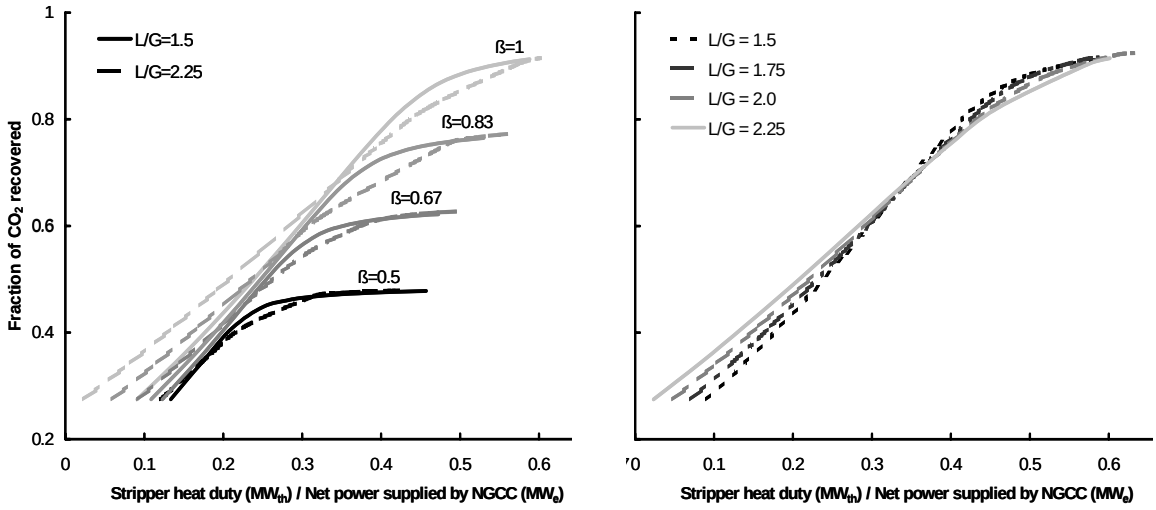

Fig. 6 a (left) CO_2 recovered as a function of the heat required to regenerate the MEA; b (right) CO_2 recoveries for different values of the bifurcation ratio (ß)

Heat and power integration

Large heat duties are associated with high CO_2 recoveries in the MEA unit, as shown in Fig. 6. This heat, necessary to regenerate the amine, is typically drawn from the steam cycle and significantly reduces the net efficiency of the power plant. Although the stripper heat duty is key in the MEA recovery unit, there are a number of other important energy requirements. They include several heating and cooling duties to meet the input conditions of liquid and gas streams, cooling requirements of the effluent gas stream of the stripper necessary to eliminate condensable components and electrical energy to move the large gas and liquid volumes. We have assessed energy savings through the heat and power integration of the overall process. To minimize utility requirements not only the amount of heat matters but also the temperature levels. Best matches between hot and cold streams within the required temperature levels produce utility savings in the most efficient manner. Heat can be recovered from the hot streams, namely the flue gas and the liquid leaving the stripper to supply the heat required to condition the feed entering the stripper to 93 °C. As a result, hot utility is not necessary and only cooling utility (150 MW_{th}) is required to cool down the inlet stream of the absorber to 40 °C.

Energy cascading within the NGCC is represented in Fig. 7. The heat from the combustion of natural gas is used to heat up the working fluid in the gas turbine system (GTS, composed by the 6 gas turbines) from 300 to 1000 °C. The GTS generates power (365 MW_e) and releases about 730 MW_{th} from 530 to 100 °C. This amount of heat plus 195 MW_{th} of additional heat supply from supplementary firing are used to generate high pressure (HP) and low pressure (LP) steam. The steam turbine generates power (229 MW_e) and releases 653 MW_{th}. In the energy cascading representation of Fig. 7b, the stripper can also be considered as a thermal machine that receives heat (285 MW_{th}) at a high temperature (107 °C) generates work (the separation work that is necessary to release the CO_2 from the amine solution and regenerate the solvent) and releases heat at a low temperature (40 °C). This “mass transfer machine” is heat integrated in the NGCC by using LP steam that is easiest to draw from the steam turbine cycle. Almost 375 MW_{th} of cooling utilities are needed to satisfy the capture system requirements, 120 MW_{th} for flue gas treatment, 150 MW_{th} for lean amine and 105 MW_{th} for the condenser. The LP steam used as heating requirement, returns to the steam cycle as condensed water at 126 ºC, there is a cooling duty saving of 256 MW_{th} of the steam cycle which is used for the capture system reducing the cooling requirements to 119 MW_{th} this corresponds to a nominal flow rate of cooling water of about 2.95 10^6 m^3 per month. This figure is above the water usage limit of the nearby Limay river that is fixed on 1.1 10^6 m^3 per month. The

NGCC already uses almost 0.6 10^6 m^3 per month. This situation implies that a cooling system is necessary to supply the cooling duty for the MEA recovery unit.

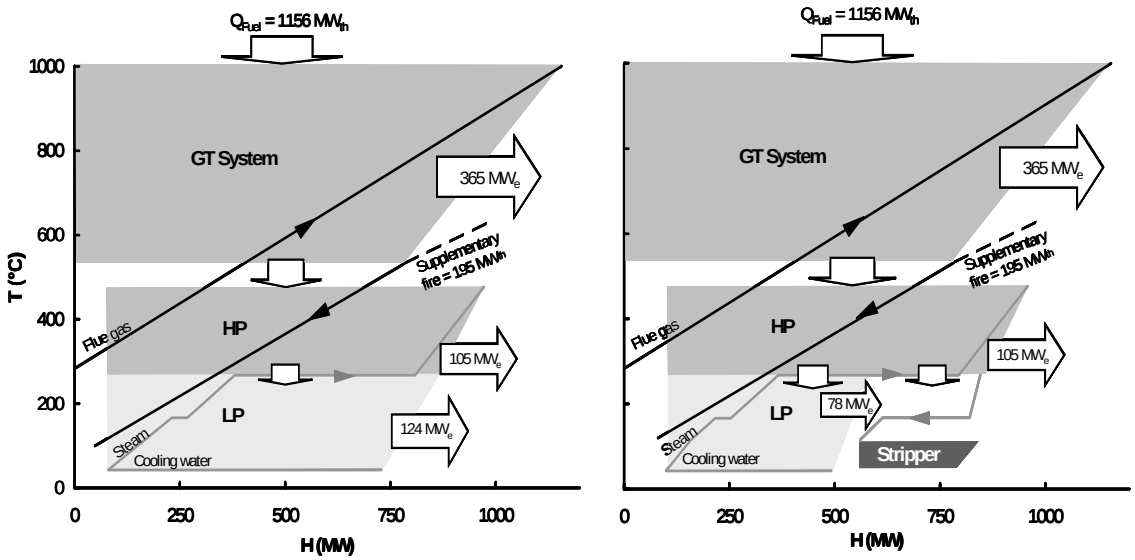

Fig. 7. a (left) Thermal energy cascading of the reference NGCC. The evolution of the flue gas, steam and the power generated by the turbines are shown; b (right) Thermal energy cascading of the reference NGCC with the heat integrated stripper

An overall picture of the design of the post-combustion capture unit is presented in Table 1 and illustrates the large heat requirements and equipment size involved for the reference NGCC, a typical newly installed generating machine.

Table 1. Key operating and equipment size results for post-combustion capture for the reference NGCC (594 MW_e, 44% efficiency)

Parameter	Units	Ref. NGCC with CO_2 capture
Net plant capacity	MW_e	515
CO_2 captured	Mton/yr	1.8
Efficiency	%	38
Reboiler heat duty	MW_{th}	285
Cooling duty, MEA recovery unit	MW_{th}	375
Power (flue gas compression + pumps)	MW_e	32
Absorber		6 columns, D=10 m, H=35 m
Stripper		2 columns, max D=7 m, H=30 m

Cost estimation

Capital and operation costs for the CO_2 post-combustion capture unit were estimated for our design. The total capital investment (C_{TCI}) was estimated based on the individual factors method of Guthrie[34-35] and is discussed for the whole process including compression-dehydration and transportation of the captured CO_2 in section 7. To apply the method, the f.o.b. purchase cost of each piece of major process equipment or process machinery was estimated. Sources of purchase costs involve local vendors and the recent compilation by Seider et al.[36] Next, we calculated the so called bare-module cost that accounts for the purchase cost of the equipment or machinery plus its installation. The total bare-module cost amounts to 146.8 10^6 US$ (2003) and is dominated by the cost of the MEA absorption/stripping unit that includes a large area for heat recovery and

temperature conditioning among the streams connecting the absorbers and the strippers. The costs of the 6 blowers that move the flue gas from the NGCC to the absorbers are also significant, accounting for almost 24% of the total bare-module cost. Total permanent investment is 205.3 10^6 US$ (2003) and represents about 34% of the total sum actually invested for the reference NGCC. Annual operating cost at 100% capacity is 5.3 10^6 US$ (2003).

PRE-COMBUSTION CAPTURE FROM THE REFERENCE POWER PLANT

Design of a steam-methane-reforming process

Steam methane reforming (SMR) was chosen for pre-combustion de-carbonization because it is a mature technology already in use in Argentina and is also the largest and generally the most economical way to make hydrogen.[30] The pre-combustion unit is composed of a SMR reactor (primary reformer) followed by a medium temperature water gas shift (WGS) reactor that converts CO and H_2O, components of the H_2-rich stream leaving the SMR reactor, to CO_2 and more H_2. A recovery unit where the H_2-rich fuel and the CO_2 are separated is also included. The SMR reactor consists of hundreds of parallel vertical tubes filled with nickel catalyst pellets within a furnace. The WGS reactor under typical industrial conditions operates close to the reaction equilibrium and it was thus modelled. The design of this unit, used to produce fuel gas to be burned in the gas turbine of a combined cycle, was undertaken to acquire information on energy requirements, equipment size, content of H_2, CH_4, CO and CO_2 of the exit stream for different process conditions. To validate our calculations, we selected the widely recognized work by Xu and Froment[37] as test case. We adopted the kinetics model reported in that reference and used the Peng Robinson equation of state modified by Huron and Vidal for physical properties estimation. A key parameter in the simulation of a SMR reactor is the global heat transfer coefficient (U) since the results are very sensitive to its value. Our simulation results showed a satisfactory agreement with those of the test case and best fitting was obtained with the correlation of U reported by Rajesh *et al.*[38] and was thus adopted.

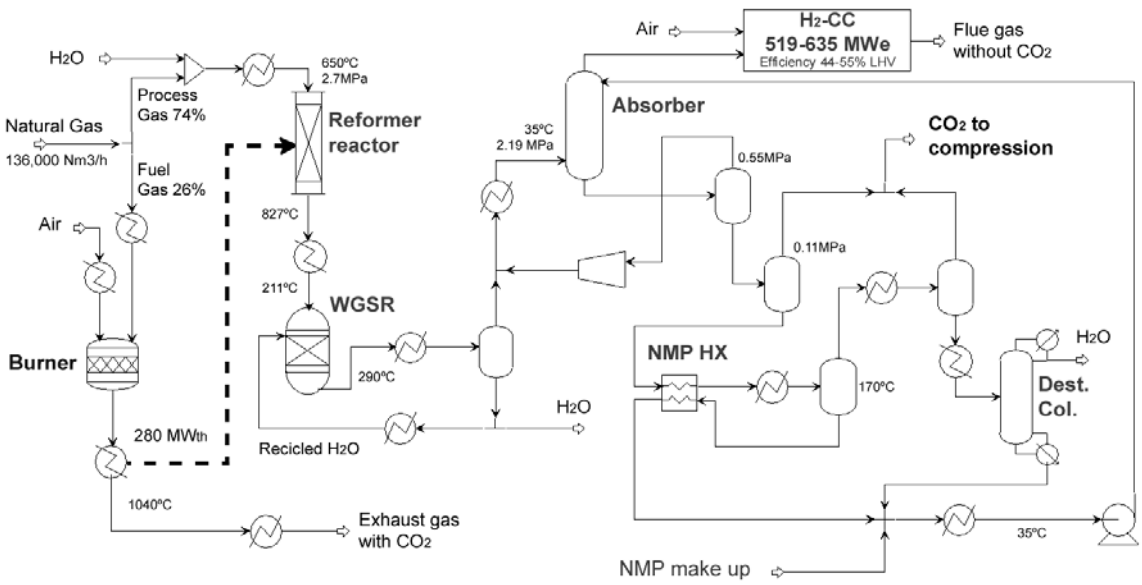

Fig. 8. Flow diagram of the pre-combustion capture unit with the main design parameters

The key variables explored to design the process were the flow rate per tube, the steam to methane (H_2O/CH_4) inlet ratio, the tube wall temperature profile and the reformer inlet pressure. Excess steam must be used to reduce by-product carbon formation[30, 38] and the H_2O/CH_4 ratio normally represents 2 to 3 times the necessary stoichiometric ratio. The natural gas feed of 136 000 $Nm^3\ h^{-1}$ was split into a fuel stream and a process stream (Fig. 8). The fuel stream is burned with air to supply the necessary heat for the endothermic reforming reaction. Due to the operating temperature

needed for the reforming, the fuel stream is higher than the theoretical amount that the reaction requires. Most of the excess heat used in the reformer is then used to pre-heat the process streams and to produce additional steam that generates electricity. The reactor was modelled on a single tube basis and the number of tubes necessary to accommodate the total input (NG process + steam) was calculated iteratively considering the specification for the split ratio between the burned gas and the process gas. This split ratio results from specifying that the heat that can be recovered from cooling down the burner outlet stream to the design temperature within the range of 1040-1100 °C be equal to the total reformer heat duty. This temperature is the needed value to allow the heat transfer between the flue gas and the process stream at the end of the reformer. Increasing and constant wall temperature profiles along the tube length were explored. The constant profile was fixed at the maximum allowed wall temperature[39] for reformer tubes of 1050 °C and the increasing profile represents a roughly logarithm type curve starting at 850 ºC and ending at 1000 ºC. Both cases implied decreasing heat fluxes profiles. Decreasing heat flux is preferable because of the higher concentrations of the reactants at the reformer inlet which implies higher reaction rate needing a higher heat flux to satisfy the strongly endothermic reaction.

Increasing the H_2O/CH_4 ratio in the reformer inlet produces two opposite effects regarding the amount of recoverable CO_2. CH_4 conversion increases with H_2O/CH_4 ratio, producing more CO_2 (or CO which is further convert it to more CO_2 in the WGSR). On the other hand, the increase in CH_4 conversion implies that more heat is required in the reformer, so the fraction of CH_4 used as fuel increases. Since the CO_2 produced in the combustion of CH_4 is not captured, this reduces the amount of recoverable CO_2. The overall balance of both effects is dominated by the increasing production of CO_2. However, because of smaller investment costs and smaller heat requirements for water evaporation, it is desirable to operate the reformer with the lowest possible H_2O/CH_4 ratio. Since the H_2-rich stream produced in the SMR unit is going to be burn as fuel in the gas turbine, the unconverted CH_4 is not a problem in that stream. Thus, H_2 production for this use allows an alternative possibility not commonly used in the high purity H_2 production. This is the introduction of additional steam after the reformer reactor to increase the conversion of CO to CO_2 in the WGS reactor. Introducing extra steam after the reformer permits obtaining the same amount of recoverable CO_2 as in the case of introducing all the steam to the reformer reactor, while reducing the number of tubes in the reformer reactor. This has the additional advantage of reducing the fraction of CH_4 used as fuel in the reformer that implies that more electric power can be produced in the gas turbine.

There are possibilities to heat integrate various hot and cold streams. Heat can be recovered from the streams leaving the SMR reactor, the WGS reactor and from cooling down the flue gas of the burner from 1100°C to 150°C. Heat is required for natural gas, air, and water preheating duties. Extra heat is obtained because the available heat from the hot streams is higher than the heat duties of the cold streams. The reformer inlet pressure was explored for values in the range 1.5 to 2.7 MPa. This parameter affects the process in two different ways. The reaction equilibrium is favoured at low pressure but the reaction rate increases at higher pressures. The operating pressure in the downstream CO_2 recovery system must be also taken into account when selecting the SMR reactor pressure. The CO_2 recovery system in pre-combustion processes is different than in post combustion essentially because of the higher CO_2 partial pressure. Under this condition it may be preferable to use physical solvents, which combine less strongly with CO_2.[40] This is the main reason why it is convenient to work with the higher possible pressure. In our case, the available pressure of the natural gas (2.7 MPa) was selected as the inlet pressure to the reformer. We evaluated four physical absorption processes (Fluor, Purisol, Rectisol and Selexol) and methyldiethanolamine (MDEA) for the recovery unit.

Total methane conversion in the range 63-95% is obtained in the SMR reactor for the range of parameters studied. This conversion increases as both the flow rate per tube and the H_2O/CH_4 ratio increase and is not significantly influenced by internal tube diameter for flow rate per tube lower than 18 kmole h^{-1}. Over this value, noticeably higher conversions are obtained for the lowest tube internal diameter. This behaviour is caused by an increasing turbulence (higher Reynolds number)

that enhances the heat transfer between the wall and the process stream and a higher pressure drop that shifts the overall reaction towards products. Product distribution and remnant CH_4 are shown for different H_2O/CH_4 ratios in Fig. 9 as functions of the feed per tube flow rate, for the internal diameter of 10.16 cm and a reformer inlet pressure of 2.7 MPa. A similar behaviour was observed for the others diameters analyzed. Mass flow rates in Fig. 9 correspond to exit conditions of the WGSR after water condensation. The distribution shows that higher H_2 (and CO_2) productions are obtained with increasing levels of steam. This is due to the role of the WGS reaction ($CO + H_2O = H_2 + CO_2$) whose equilibrium is shifted towards product formation with increasing steam levels. This fact is confirmed with the opposite behaviour of CO. H_2 mass flow rate is the least sensitive with respect to flow rate per tube showing an increase of about 20% with an increase from 10 to 27 kmole h^{-1} per tube. More drastic variations can be seen for remnant CH_4 with a decrease of about 87% for the same range. For the smallest diameter, not only higher pressure drops occur but also the trend of this variable noticeably increases with increments of the rate per tube. Total tube number varies within the range 400-1000 depending on the flow rate per tube and the H_2O/CH_4 ratio. Typically a radiation zone of a primary reformer features a tube bundle comprising different amount of tubes according to the type of furnace, namely 600-1000 (up-fired); 100-150 (lateral fired); or 200-300 (down-fired). These values constitute important bounds for the final design of the SMR reactor.

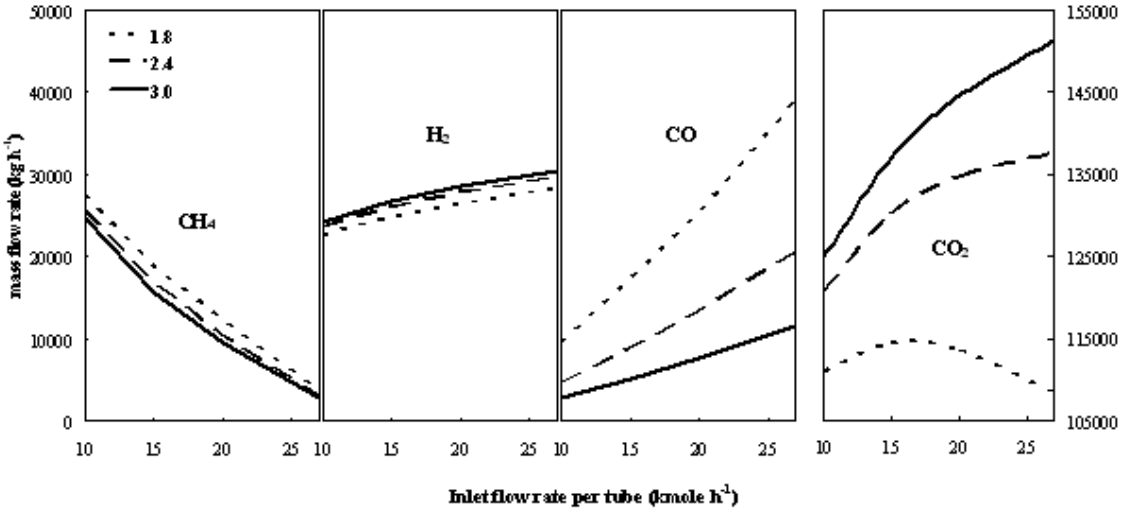

Fig. 9. Mass flow rate of reactant and products leaving the WGS reactor as functions of the inlet rate per tube.

For the CO_2 recovery unit, the five separation processes considered were evaluated for a specific case of the reaction unit, defined by a flow rate per tube of 27 kmole h^{-1}, a tube diameter of 12.56 cm and a value of 2.4 for the inlet H_2O/CH_4 ratio. A flow rate of 18 000 kmole h^{-1} with a mole composition of H_2 (77.3%), CO_2 (16.6%), CO (3.6%), CH_4 (1.75%) and H_2O (0.63%) was input to the CO_2 recovery unit at 50 °C and 2.14 MPa. All the separation processes showed acceptable performances in regard to CO_2 recovery being the MDEA process that with the largest heating and cooling duties. The energy required by the Fluor, Selexol and Purisol processes was intermediate but the Fluor process operates at higher pressure (~ 100 MPa) than the others. The Rectisol process exhibited the lowest heat requirements but a refrigeration cycle is necessary by the low temperatures involved. A more detailed analyses was undertook between Purisol and Selexol process which revealed that Purisol process has less energy requirements when the goal is CO_2 bulk recovery. Because of its relatively lower equipment cost regarding its operating pressure and temperature levels, the Purisol process was adopted for our final design.

We assumed that the exiting H_2-rich fuel from the CO_2 recovery unit can be burned in a new gas turbine belonging to a CC with a hypothetical efficiency in the range 44-55%. The extra heat available from the SMR unit can be also used in the HSRG of the hypothetical CC. Based on these

contributions and the assumed efficiency, an achievable power was estimated within the range 519-635 MW_e. The results for the overall pre-combustion unit are set forth in Table 2.

Table 2. Key operating and equipment size results for pre-combustion unit as compared with the Reference NGCC (594 MWe and 44% efficiency)

Parameter	Units	H_2-rich fuelled CC
Net plant capacity	MW_e	519-635
CO_2 captured	Mton yr^{-1}	1.01
Efficiency	%	38.4 - 47.0*
Number of tubes (14 cm, 13 m) of SMR		464
Extra heat available from the SMR unit	MW_{th}	32
Power (pump, compressor) CO_2 recovery unit	MW_e	6.4
Cooling water for the Purisol process	MW_{th}	66

* Referred to 136,000 Nm^3/h of natural gas input (process + fuel necessary to satisfy the reformer heating duty).

Cost estimation

Capital and operation costs for the CO_2 pre-combustion capture unit were estimated using the same method as discussed in section 4.3. The total bare-module cost amounts to 328.6 10^6 US$ (2003) and is dominated by the cost of the SMR unit and the large area for heat exchangers. From the 135 heat exchangers required in the pre-combustion unit, 122 belong to the heat recovery between hot and cold streams of the NMP solvent used by the Purisol process. These heat exchangers account for 92% of the overall bare-module cost of the heat exchangers (155.6 10^6 US$). The mass transfer equipment (absorber + distillation column) accounts for less than 2% of the total bare-module cost. Total permanent investment amounts to 453.7 10^6 US$ (2003) and represents about 76% of the total sum actually invested for the reference NGCC. Annual O&M cost at 100% capacity is 11.1 10^6 US$ (2003).

TRANSMISSION AND INJECTION OF CO_2

The two possibilities that were considered for routing the hypothetical pipeline between the reference power plant and the prospective storage site are shown in Fig. 3. They are 71.8 km and 96.8 km contour lines that were selected on the basis of existing roads and/or natural gas pipelines. The conceptual design of the pipeline taking into account the characteristics of the local terrain, particularly its elevation, is relevant because static-head variations (increases as flowing downhill or decreases when flowing uphill) influence the temperature profile in the pipeline, which in terms noticeably affects the flowing properties of CO_2.[25] Apart from the terrain characteristics, the cost of the pipeline is a function of pipe diameter, pipeline inlet pressure, booster pumps and pressure regulators. The requirements concerning compression and dehydration of the CO_2 streams leaving the capture systems to meet the specifications for pipeline transmission were also assessed.

All calculations were done using the Peng-Robinson model for physical properties estimation. The overall error in the estimation of pure CO_2 compressibility factors in the range 0-1000 °C, 0.1-50 MPa with respect to reference results[41] was 0.08%. The performance of the simulator for CO_2 pipeline design was validated on the basis of the example presented by Farris[25] concerning a 386 km pipeline transporting 98.3 m^3/s of pure CO_2 in a 16-in pipe.

Conditioning of the CO_2 streams leaving the post and pre-combustion units

The captured CO_2 streams must be dehydrated and compressed before being input to the pipeline. A simplified scheme of the compression-dehydration unit that employs TEG as dehydrating agent is shown in Fig. 10.

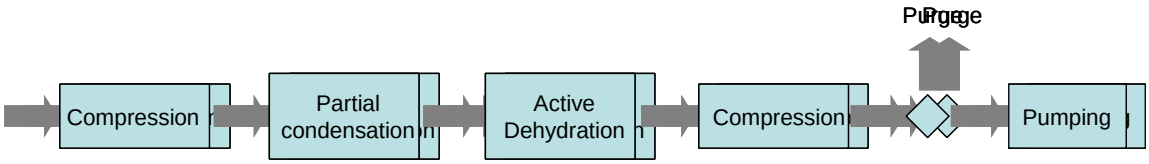

Fig. 10. Simplified scheme of a CO_2 conditioning unit

Partial condensation for the streams leaving post and pre-combustion capture was assessed for the operating ranges: -30 °C to 30 °C and 0.5-4.5 MPa. Our results showed that at 30 °C and pressures > 3 MPa the amount of H_2O condensed does not increases with increasing pressures while the CO_2 content in the condensed water is always below 0.1%. The definite partial condensation conditions are determined by the optimal operating conditions of the TEG dehydration column that must assure an exit stream with a maximum H_2O content in the range 60-100 mg/m^3. Different conditions involving operating pressure, TEG flow rate and number of stages were explored for active dehydration. Simulation results show that increasing pressure up to 3 MPa favours water absorption in TEG. Beyond this limit, the amount of water absorbed decreases noticeably with increasing pressures. The final design for the dehydration column consists of 10 theoretical stages, average operating conditions of 30 °C and 3 MPa and a flow rate of TEG of 5 kmol/h for the pre-combustion stream and of 9 kmol/h for the post-combustion stream. CO_2 concentration of the dehydrated streams is ~ 98% for the pre-combustion stream and almost 100% for the post-combustion stream. After dehydration, the diluted TEG stream leaving the absorption column is fed to a distillation column where TEG is recovered and recycled.

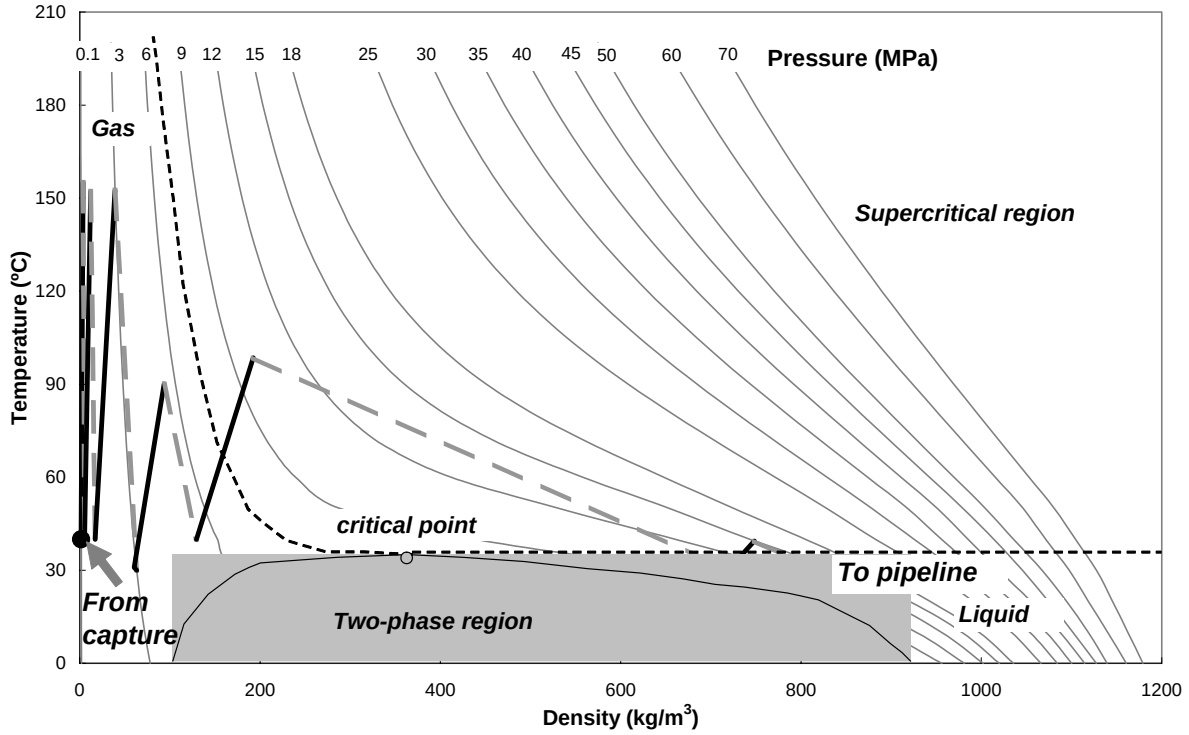

Fig. 11. The CO_2 stream leaving the capture systems is subjected to a series of compression and cooling steps to reach the conditions adequate for pipeline transmission. These steps are plotted as a segmented line on a Temperature-Density diagram of pure CO_2. The straight black line indicates compression; the dashed grey line indicates cooling.

The compression from 0.1 MPa up to 3 MPa has the highest power requirement of the CO_2 conditioning unit (~ 78% of the total) and it also has the largest investment cost. A radial 3-step compressor with a total compression ratio of 3.21 was selected for this task. Air cooled heat exchangers are used after the first 2 compression steps to cooled down the stream to ~ 40 °C and a water cooled heat exchanger is used after the final compression step. Condensed water is separated from the CO_2 stream after each cooling step. Approximately 62% of the water contained in the pre-

combustion stream and 97% of that contained in the post-combustion stream is condensed before entering the active dehydration column. After active dehydration, the pressure is increased high enough to condense CO_2 at a reasonable cost.[42] This second compression accounts for 20% of the total power of the CO_2 conditioning unit. The final pressure increase is done by means of a high pressure pump after having vented the non-condensable gases. Fig. 14 depicts the compression steps used in our study. The first 3-step compressor increases the pressure from 0.1 MPa to 3 MPa. After dehydration, the 2-step compressor increases the pressure from 3 MPa to 10 MPa bringing the CO_2 to the supercritical region. After cooling, CO_2 is liquid and is pumped up to almost 15 MPa, meeting pipeline operating conditions.

Design of the transportation and injection of CO_2 into depleted gas field

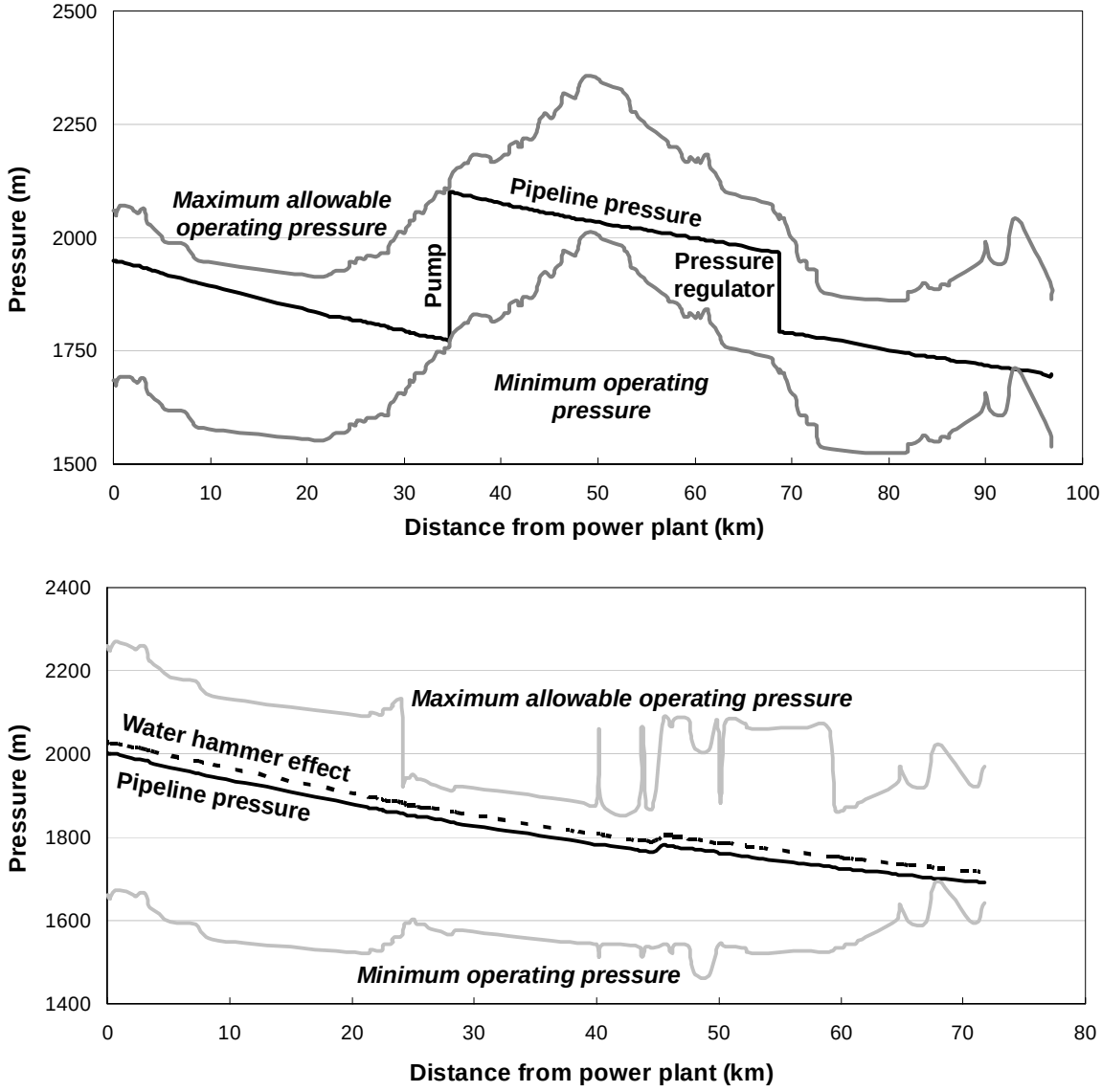

Fig. 12. Conceptual design of a CO_2 pipeline. a) Top: longest route, 12-in nominal diameter, 5.6 mm thick; b) shortest route, 12-in nominal diameter, 5.6 or 6.4 mm thick.

The two alternative routes were selected using the best available topographic maps for the area (scale of 1:100,000; obtained from the Military Geographic Institute of Argentina). The line profile elevation for each hypothetical route was developed on the basis of these maps. A series of five nominal pipe diameters of 10, 12, 14, 16 and 18 inches were explored looking for the best pipeline design. At least a pair of alternative thickness values was considered for each nominal pipe diameter.

The CO_2 pipeline was designed by calculating the pressure drop from the energy balance equation. The requirement of booster pumps and pressure regulators was evaluated from the pressure drop

across the pipeline using the minimum pressure and the maximum allowable operating pressure (MAOP) as design limits. Both pressure limits depend primarily on the terrain elevation and the MAOP depends also on the pipeline thickness. The minimum pressure limit includes a safety factor of ~20% above the critical pressure of each transported stream. The design approach is illustrated in Fig. 12a for one of the cases analysed that concerns the post-combustion stream and the longest route. When the pressure drop at a certain point is such that the pipeline pressure is close to the minimum pressure, a booster pump is needed. On the other hand, if the pipeline pressure is close to the MAOP, a regulator station is necessary. Required theoretical pump power is computed directly from enthalpy values at suction and discharge conditions assuming constant entropy**¡Error! No se encuentra el origen de la referencia.**. For many of the designs considered the inlet pipeline pressure was sufficient and no booster pump was required. This is illustrated in Fig. 12b where it can be seen that the CO_2 is transported along the 71.8 km shortest route with no intermediate booster pump or pressure regulator. Fig. 12b corresponds to a pipeline of 12-in nominal diameter and belongs to the design finally adopted for the post-combustion stream. Two values of pipe thickness of 5.2 and 6.4 mm were employed. The switches between these values are observed in the discontinuities of the MAOP profile. The 6.4 mm pipe is used in those occasions that the pipeline operating pressure exceeds the MAOP belonging to the 5.3 mm pipe. This type of option not requiring extra power was finally selected for each capture stream because of its simplicity and most importantly because extra power required at isolated intermediate sites would certainly involve extra CO_2 emissions from heat engines. For the final design, a safety factor was included to account for potential water hammer effects because of sudden flow stoppage (Fig. 12b). The characteristics of the compression-dehydration unit and the CO_2 pipeline for the pre and post-combustion streams are summarised in Table 3.

Table 3. Key operating and equipment size results for compression-dehydration and CO_2 pipelines for the pre and post-combustion streams.

Parameter	Units	Pre-combustion	Post-combustion
CO_2 flow rate to injection	ton h^{-1}	126.35	226.51
Power 3-step compressor (0.1-3 MPa)	MW_e	10.8	19.7
Power 2-step compressor (3-5.4 MPa)	MW_e	2.9	5.0
Power pump (5.4-14/15 MPa)	MW_e	0.3	0.6
TEG dehydration column	-	D=1.8m, H=11.5 m	D=2.5 m, H=11.5m
TEG recovery distillation column	-	D=0.46m, H=1.9 m	D=0.49m, H=1.9 m
Cooling water	MW_{th}	13.3	24.2
Heat reboiler TEG recovery column	MW_{th}	0.321	0.424
71.8 km pipeline			
nominal diameter / thickness	inches / mm	12 / 5.6	12 / 5.6-6.4
weight	ton	3154	3380
Power for injection	MW_e	0.6	1.2

COST ESTIMATION

Capital and operation costs for the compression-dehydration unit were estimated using the same method as discussed in section 4.3 The total bare-module costs of the compression and dehydration units are: 41.3 10^6 US$ (2003) for the pre-combustion system and 66.0 10^6 US$ (2003) for the post-combustion system. The costs of the compressors are the dominant feature of this unit accounting for almost 89% of the total bare-module cost. Total permanent investment for the pre and post-

combustion compression-dehydration units are 41.3 and 66.0 10^6 US$ (2003) respectively while annual O&M costs are 1.2 and 1.8 10^6 US$ (2003) respectively.

The cost corresponding to the CO_2 pipeline was calculated according to the estimation scheme used by one of the natural gas transmission companies of Argentina. It is based on the cost per mass of steel and includes percentage estimates of other materials, construction, supervision and right-of-ways. The estimated cost of the installed pipeline are, for a 12-inch nominal pipe, within the range 12-14.5 US$ (2003)/in-m that is the standard manner of representing this cost in the estimation scheme. These figures implies an installed cost of 12.5 10^6 US$ (2003) for the pre-combustion system and 11.7 10^6 US$ (2003) for the post-combustion system. Total permanent investment for the pre and post-combustion pipeline are 13.8 and 14.6 10^6 US$ (2003) respectively while annual O&M cost is ~0.4 10^6 US$ (2003) for each pipeline.

Cost effectiveness of CO_2 capture and storage

Capital and operation costs were estimated for our systems after having completed the conceptual design. For each CO_2 capture option, the cost estimates include: *i*) capture unit, *ii*) compression-dehydration unit and *iii*) pipeline and injection. A cash flow projection was used to determine the additional cost of electricity production when CO_2 capture and storage is added to the reference NGCC. Finally the overall electricity production costs are compared with the cost of electricity production from nuclear power. These estimates are summarised below.

Overall costs of CO_2 capture and storage for the reference power plant

Costs for our case study were estimated for year 2003 since the economical and financial crisis of Argentina that occurred at the end of 2001 provoked among other things drastic changes in the currency exchange rate. The peso ($) exchange rate with the US dollar (US$) was maintained fixed at 1 $/US$ until 2001 when the economic situation worsened. Government efforts to stabilise the banking system and to restore economic growth proved inadequate in the face of the mounting economic problems. The peso's peg to the dollar was abandoned in January 2002, and the peso was floated in February; the exchange rate plunged to almost 0.25 $/US$ and GDP fell by 10.9% in 2002. However, by mid-year the economy had stabilised at a lower level. GDP expanded by more than 8% in 2003 and again in 2004, with exchange rate rather stabilised at an average of 0.33 $/US$. This exchange rate value was used for our cost estimations.

As explained in sections 4.3, 5.2 and 6.3, the total bare-module costs (C_{TBM}) of the major equipment of three units of the capture and storage system were estimated based on the individual factors method of Guthrie.[34-35] For determining the total permanent investment (C_{TPI}) of each capture option additional charges were applied using a standardized factored-cost method. The results on total permanent investment are summarised in Table 4. The cost for both systems is comparable with the original investment done to install the reference NGCC: The pre-combustion capture system amounts to 85% of the investment for the NGCC while that of the post-combustion system is equivalent to 48%. It is clearly evident that the cost of the capture unit is the dominant factor of the total permanent investment of both options. It accounts for 89% of the total investment for the pre-combustion system and for 72% of the total investment of the post-combustion system. The investment for the compression-dehydration units is also significant, accounting for 8% and 23% of the total investment for the pre and post-combustion systems respectively. The ~72 km pipeline constitutes the lesser part of the investment.

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Table 4. Total permanent investment for the pre-combustion and the post-combustion systems (Million US$ dollars, 2003).

	CO_2 pre-combustion capture system				CO_2 post-combustion capture system			
	Capture	**Comp-dehyd.**	**Pipeline**	**Total**	**Capture**	**Comp-dehyd.**	**Pipeline**	**Total**
C_{TBM}: bare module	**328.6**	**29.8**	**11.7**	**370.2**	**147.5**	**47.6**	**12.5**	**207.5**
C_{site} + C_{serv}: site & services	16.4	1.5	0.0	**17.9**	7.4	2.4	0.0	**9.8**
C_{alloc}: up-grade facilities	1.0	0.2	0.0	**1.2**	1.7	0.4	0.0	**2.1**
Direct permanent investment	**346.1**	**31.5**	**13.7**	**389.3**	**156.6**	**50.3**	**12.5**	**219.4**
C_{cont}: contingencies	51.9	4.7	1.8	**58.4**	23.5	7.5	1.9	**32.9**
Total depreciable capital	**398.0**	**36.3**	**13.5**	**447.7**	**180.1**	**57.9**	**14.3**	**252.3**
C_{land}: land	8.0	0.7	0.0	**8.7**	3.6	1.2	0.0	**4.8**
$C_{royalties}$: royaties	8.0	0.7	0.0	**9.0**	3.6	1.2	0.0	**4.8**
C_{start}: start-up	39.8	3.6	0.3	**43.4**	18.0	5.8	0.3	**24.1**
Total permanent investment	**453.7**	**41.3**	**13.8**	**508.8**	**205.3**	**66.0**	**14.6**	**285.9**

Table 5. Annual operating and maintenance cost for the pre-combustion and the post-combustion systems (Million US$ dollars, 2003).

	CO_2 pre-combustion capture system				CO_2 post-combustion capture system			
	Capture	**Comp-dehyd.**	**Pipeline**	**Total**	**Capture**	**Comp-dehyd.**	**Pipeline**	**Total**
Feedstock								
Solvent reposition.	4.96	0.05	0.00	**5.01**	0.07	0.05	0.00	**0.11**
Process water	0.24	0.00	0.00	**0.24**	0.09	0.00	0.00	**0.09**
Labour cost								
Direct labour (4 shifts)	0.14	0.07	0.07	**0.29**	0.11	0.07	0.07	**0.25**
Supervision	0.03	0.01	0.01	**0.06**	0.02	0.01	0.01	**0.05**
Administration	0.09	0.04	0.04	**0.17**	0.06	0.04	0.04	**0.15**
Maintenance								
Wages & benefits	2.44	0.44	0.16	**3.04**	2.17	0.70	0.17	**3.04**
Salaries & benefits	0.61	0.11	0.04	**0.77**	0.54	0.17	0.04	**0.76**
Materials and services	2.44	0.44	0.04	**2.92**	2.17	0.70	0.04	**2.92**
Maintenance overhead	0.12	0.02	0.00	**0.15**	0.11	0.03	0.00	**0.15**
Total	**11.07**	**1.18**	**0.39**	**12.64**	**5.35**	**1.78**	**0.39**	**7.52**

Annual O&M costs include the cost of solvent reposition for the different absorption processes, process water, labour and operation and maintenance costs (Table 5). Direct labour accounted for 4 shifts of 7-8 operators for the post or pre-combustion options respectively. The remaining labour cost components were estimated as percentages of the direct cost labour costs. The maintenance costs were estimated as percentages of the total depreciable capital. The operating cost of the pre-combustion system is 68% higher than that of the post-combustion system. This difference is mainly due to the cost of reposition of the NMP solvent for the Purisol process as compared to the reposition cost of MEA. The O&M costs are also dominated by the costs of the capture units that account for ~88% of the O&M cost of the pre-combustion system and 71% of the O&M cost of the post-combustion system.

The cash flow projection was used to determine the additional cost of electricity production that is incurred when a capture system is added to the reference NGCC. We considered the cash flows of the capture systems throughout the lifetime of the power plant and calculated the net present value (NPV) of these cash flows. The cash flows include *i*) sales revenue from selling electricity; *ii*) O&M costs; *iii*) capital expenditures; *iv*) working capital; and *v*) depreciation. The additional production cost per MWh of electricity was the variable of the cash flow projection and was calculated by setting the NPV of the added capture system to zero. This was achieved by varying the MWh price until the sales revenue balance the cost over the whole lifetime considered. The main parameters used for the clash flow are summarised as follows. Lifetime period: 25 years with 2 additional years for construction. Linear depreciation assumed and class lives of 5, 10 and 15 years. Interest rates: 10 and 15%. Income taxes of 21% are applied in Argentina to electricity generators. A null value of income taxes was also used for comparative purposes because some of the studies on CO_2 capture have been performed under this assumption. The results obtained are shown in Table 6 where they are also compared with previously reported data.[43-48] The additional costs obtained are within the range of the selected references and highlight the significant efforts that are needed to carry out a CO_2 capture and storage project.

Table 6. Additional electricity cost in US$ (2003)/MWh

	Income tax = 21%	No income tax	Previously reported data
Pre-combustion capture	20.5 – 30.5	18.8 – 26.1	9.7 – 31
Post-combustion capture	12.1 – 17.9	11.1 – 15.4	8.4 – 18
References: 43-48			

The reference NGCC had an electricity generation cost of 8.5 US$/MWh in 2003. This relatively low value is related to the two aspects namely, the relative high thermal efficiency of the reference NGCC as compared to other power plants of the WEM and the low cost of the natural gas for this particular NGCC that is located close to the gas field. The lowest electricity generation cost (7.2 US$/MWh) in the Argentinean WEM, corresponded in 2003 to another NGCC using relatively cheap natural gas from nearby gas fields. However, this NGCC is located in the Northwest region of the WEM where CO_2 storage has been classified as non-prospective.

In the most favourable cases, taking into account the income taxes of 21% that are paid in Argentina, the cost of electricity for the reference NGCC with CO_2 capture and storage would be 20.6 US$/MWh for post-combustion capture and 29 US$/MWh for pre-combustion capture. For this last case, the electricity generation cost of the reference NGCC (without CO_2 capture) was applied to the hypothetical H_2-rich-CC that

would be used in pre-combustion option. Although this assumption has inherent uncertainties, it was considered adequate for comparative purposes.

CONCLUSIONS

This work has focused on the reference power plant to assess the implications of CO_2 capture from recently installed NGCCs in Argentina. Although it is possible to retrieve technology and cost information from the open literature and extrapolate these data to the local circumstances, we aimed at establishing a framework for quantifying the key variables involved in post and pre-combustion CO_2 capture systems from thermal power plants.

Our results on heat duties, equipment size and cost estimates are descriptive of the extensive requirements associated to this type of projects. The SMR-based pre-combustion system exhibited additional electricity costs ~70% higher than those corresponding to the MEA-based post-combustion system. It is worth contrasting this result with some considerations on post combustion capture that sometimes present this option as an un-innovative end-of-pipe approach that does not allow the synergies offered by pre-combustion capture. In principle this is not the case if the reboiler of the MEA regenerating stripper is efficiently heat-integrated into the power cycle. Furthermore, pre-combustion capture could be an interesting option if fuel decarbonisation is an inherent part of the technology such as in IGCC or if hydrogen is also a desired product but decarbonising a rather “clean” fuel such as natural gas with the main purpose of removing CO_2 does not appear as an attractive option for thermal power plants.

The large heat exchanger area that is required in both capture options is also worth mentioning. The focus on heat requirement for MEA-based post-combustion systems has always been on the reboiler of the stripper because of the significant amount of steam that is drawn from the HRSG. While this is true, the large heat exchange occurring between the hot and cold streams that interconnect the MEA-absorber with the stripper has often been disregarded. Although this heat exchange does not require extra steam, it does require cooling water and also a significant amount of heat is recovered between process streams. A rather large area is required to allow these heat exchanges and its cost constitutes ~33% of the total bare-module cost of the post-combustion unit. A similar situation occurs in the CO_2 recovery unit of the pre-combustion option where the associated heat exchangers account for ~42% of the total bare-module cost of the pre-combustion unit.

Total permanent investment for both systems is comparable with the original investment done to install the reference NGCC. The pre-combustion capture system amounts to 85% of the investment for the NGCC while that of the post-combustion system is equivalent to 48%. Our results also confirm that the total investment and O&M costs of the capture unit constitute the dominant factor of the overall costs of both options. The costs associated to the compression-dehydration units are also significant while the costs of the pipeline are much smaller.

Although our case-study has been developed based on the reference NGCC, the results on additional electricity costs can be extrapolated to practically all gas turbines and NGCC in Argentina since they burn natural gas with practically the same excess air. The suggested extrapolations are reliable because these conditions imply the same CO_2 levels in flue gas for post-combustion capture while for the analysed pre-combustion system only the fuel type matters.

It is worth mentioning that this study is about reducing emissions of CO_2 from NGCC that is a technology with relatively low emissions as compared with other power plants

burning other fossil fuels. Moreover, NGCC constitutes a standard goal for those countries where fuel switching is an attractive option. This is not the case for Argentina where electricity is mainly generated from natural gas, hydro and nuclear power. If CO_2 from large stationary sources is to be captured and stored in our country, NGCC would most probably not be the first candidate. Storage of CO_2 that is now vented from natural gas processing in highly prospective basins constitutes a much earlier opportunity that may become attractive to private actors.

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