International Journal of Greenhouse Gas Control xxx (2012) xxx–xxx

Contents lists available at SciVerse ScienceDirect

International Journal of Greenhouse Gas Control

journal homepage: www.elsevier.com/locate/ijggc

Optimisation of lean vapour compression (LVC) as an option for post-combustion CO_2 capture: Net present value maximisation

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ARTICLE INFO

Article history:
Received 13 March 2012
Received in revised form 5 September 2012
Accepted 7 September 2012
Available online xxx

Keywords:
CO_2 capture
Lean vapour compression
Techno-economic evaluation
Financial optimisation

ABSTRACT

Many process schemes have been proposed in literature to decrease the energy demand of amine based carbon dioxide processes. These process schemes are generally analysed in terms of energy demand savings and compared to a common baseline based on the solvent monoethanolamine (MEA). In this work, the application of one of these process schemes (lean vapour compression or LVC) has been evaluated and optimised based on maximising the net present value (NPV) of the process scheme savings (including capital investment), rather than minimising energy demand in the form of equivalent work. Two scenarios have been analysed. In the first scenario, the capture plant was fully adapted to the effect of LVC. In the second scenario, LVC is retrofitted to a basic capture plant design. For both scenarios the net present value (expressed in M€) of the process scheme over the whole plant life was calculated as a function of the LVC operating conditions. It was found that the NPV of the LVC process scheme is always positive and attractive from a financial point of view. The first scenario has been identified as the most attractive scenario for LVC application. Although the extent of the savings depends on design conditions and financial assumptions, this approach shows that the optimisation based on minimising equivalent work does not necessarily match the optimisation based on maximising net present value.

1. Introduction

Amine CO_2 absorption systems are considered as the most suitable technologies to capture carbon dioxide from flue gases in the power sector. The reasons behind this technological choice are, among others, the availability and reliability of the technology (already proven at smaller scale) and the possibility of retrofitting existing power plants (Gibbins and Chalmers, 2011; Wang et al., 2011). One major hurdle in the implementation of these technologies at large scale is the cost of carbon dioxide capture, which prevents utility companies from making the necessary investments. This cost varies per capture process and per power production process (Abu-Zahra et al., 2007; Davison and Thambimuthu, 2009; Finkenrath, 2011; Rubin et al., 2007). In the case of bituminous coal fired power plants, the cost of electricity can increase from 46% to 77% (Finkenrath, 2011). The reduction of this huge impact in electricity cost will be a condition for the application of CO_2 capture as a carbon abatement technology. In order to reduce costs, one of the main focuses of CO_2 capture research is the reduction of its energy demand. There are two main research lines to reduce energy demand: improvements in solvent formulation or improvement in the capture process for an existing solvent formulation. This research is concerned with the latter option.

There are different process schemes proposed in the literature to decrease the operating costs of the CO_2 capture process. Normally, new process schemes are evaluated and compared to a conventional amine scrubbing process based on monoethanolamine (MEA), which has become a common baseline for capture processes (Franco et al., 2010). These schemes can be classified in three groups, depending on the type of modification introduced with respect to the baseline (Le Moullec and Kanniche, 2010):

1. Operating modifications without addition of new unit operation to the process.
2. Minor process modifications with limited number of new unit operations added to the process.
3. Major process modifications. In this category fall the more complex changes to the capture process, such as, intermediate stripper heating (which requires full modification of the stripper), integration of stripper and compressor, etc.

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http://dx.doi.org/10.1016/j.ijggc.2012.09.007

Nomenclature

LMTD	logarithmic mean temperature difference (°C)
LRHX	lean-rich heat exchanger
LVC	lean vapour compression
MEA	monoethanolamine
NPV	net present value
Q_r	reboiler duty (GJ_{th})
W_p	work to drive process pumps (GJ)
W_{bc}	work to drive plant blower and LVC compressor (GJ)
W_{eq}	equivalent work necessary to separate CO_2 (GJ)
CF	cash flow in a given project year (M€)
ΔTC	difference in total equipment cost between the reference case (at 1.8 bar), which is 2.72 M€ and any other given case
ΔDepreciation	depreciation difference between the reference case and any other given case

Greek symbols

α	Turbine power loss to reboiler duty ratio (−)

Subscripts

i	i-th year in cash flow calculations
HEX	heat exchangers
COMP	compressor
VESSEL	vessel

This work is focused on the second group and, more specifically on the LVC scheme. This scheme includes flashing the lean hot stream that leaves the stripper in a conventional amine scrubbing process, re-compressing the vapour formed and re-injecting it at the stripper's base (definitions and description of a conventional MEA process are further explained in the following sections in this paper). This simple arrangement allows for partially recovering the sensible heat of the hot lean stream in the form of latent heat and, it is anticipated, that has potential to reduce energy demand without increasing greatly process complexity.

There are various studies in the literature that have addressed this process scheme and investigated its effect in capture performance in various ways. Van Wagener and Rochelle (2011) included lean vapour compression in their study on stripper configurations. Their work is based on the minimisation of equivalent work in the stripper and indicates the benefits of stripping at higher pressures (above 6 bar). However, their evaluation is only concerned with the stripper optimisation and the absorber is left out of scope. Le Moullec and Kanniche (2010, 2011) have also investigated lean vapour compression in a technical evaluation of process options. This study includes integration aspects with the host power plant and shows a reduction in plant efficiency penalty from 11.95% (penalty points) to 11.2% (penalty points) for LVC, which is the highest reduction given in their work when only minor modifications to the process are considered. Nevertheless, extra investments required for the application of LVC to the conventional MEA process are not taken into account in this study. The evaluation method is only based on the minimisation of the plant efficiency penalty. Karimi et al. (2011) have also included this option in a techno-economic review of process capture options. In this work, LVC is also addressed and the authors conclude that it will reduce equivalent work by 9.37% with respect to a standard MEA plant at the expense of increasing capital investment by 2.78%. In this case, the optimal operating conditions of the LVC are initially found by minimizing energy demand and capital investments are considered in a second stage for an energy optimized design.

In this study, the effect of the application of LVC on a conventional MEA capture process has also been investigated. The focus in this paper is on the design modifications, associated capital investments and optimisation of operating conditions. As opposed to the studies above mentioned, the method applied here consists in optimising process conditions based on financial analysis rather than energy analysis.

The study is restricted to a conventional design of a MEA plant that could be easily implemented in relatively short term. Therefore, high pressure desorption has been left outside of this paper's scope. Two different scenarios are analysed: optimised design of a standard MEA plant with LVC and retrofit of LVC to an existing standard MEA plant.

For each scenario, operating conditions are evaluated and optimised based on maximising the net present value (NPV) of the process modification.

This paper is organised as follows. In Section 2, the methodology is described. The reference capture process, LVC design boundary conditions and modelling technique are addressed in this section. Our findings, which include among others the financial benefit of implementing LVC regardless the scenario considered, are summarized in Section 3. For the two scenarios considered, the energy demand of the process and the net present value of the process, are discussed and compared to other literature sources in Section 4. Finally, our conclusions are summarised in Section 5.

2. Methodology

2.1. Reference case plant description

The scenarios analysed in this work are compared to a reference case, in which a 30 %w/w MEA solution is used to capture 90% of the CO_2 present in a given flue gas stream. The reference case, here referred to as a standard MEA plant, is schematically depicted in Fig. 1. The plant processes 254 kg/s of flue gas containing 13.75 mol% CO_2. This stream is a split stream of the flue gas generated by a 1070 MW Advanced Supercritical coal fired power plant and is 250 MWe equiv. (Huizeling and Weijde, 2011). The flue-gas is fed to the absorber (C-01), where it is brought in contact with the solvent, which flows downwards over a packed bed. In order to minimise MEA losses, the CO_2 lean flue-gas is washed and cooled in two washing beds at the top of the absorber. The resulting scrubbing liquid is cooled with cooling water in a separate cooler (E-05). The CO_2 rich solvent is pumped to the stripper (C-02) via a counter current lean-rich heat exchanger (LRHX) (E-03). In the stripper, the chemically bound CO_2 is removed from the solvent in a packed bed at the expense of reboiler heat. The reboiler (E-01) receives all solvent leaving the bed and partially vaporizes the stream before feeding it to the stripper sump. Most of the water in the CO_2 stream leaving the stripper is condensed (E-02) and fed to a washing section at the top of the stripper. The lean liquid is pumped through the lean-rich exchanger and cooled to 40 °C in a cooler (E-04) before being sent back to the absorber.

2.2. Lean vapour compression (LVC): design boundaries and scenarios

Application of LVC to the process described in the previous section requires additional equipment. The new LVC lines are highlighted in Fig. 1. In this case, the hot lean solvent is flashed in an additional vessel (V-02). The resulting vapour is compressed (K-02) to a pressure slightly higher than the stripper bottom pressure and fed to the stripper base. Additional pump work is required for the lean pump (P-01) to recover the pressure loss in the lean stream.

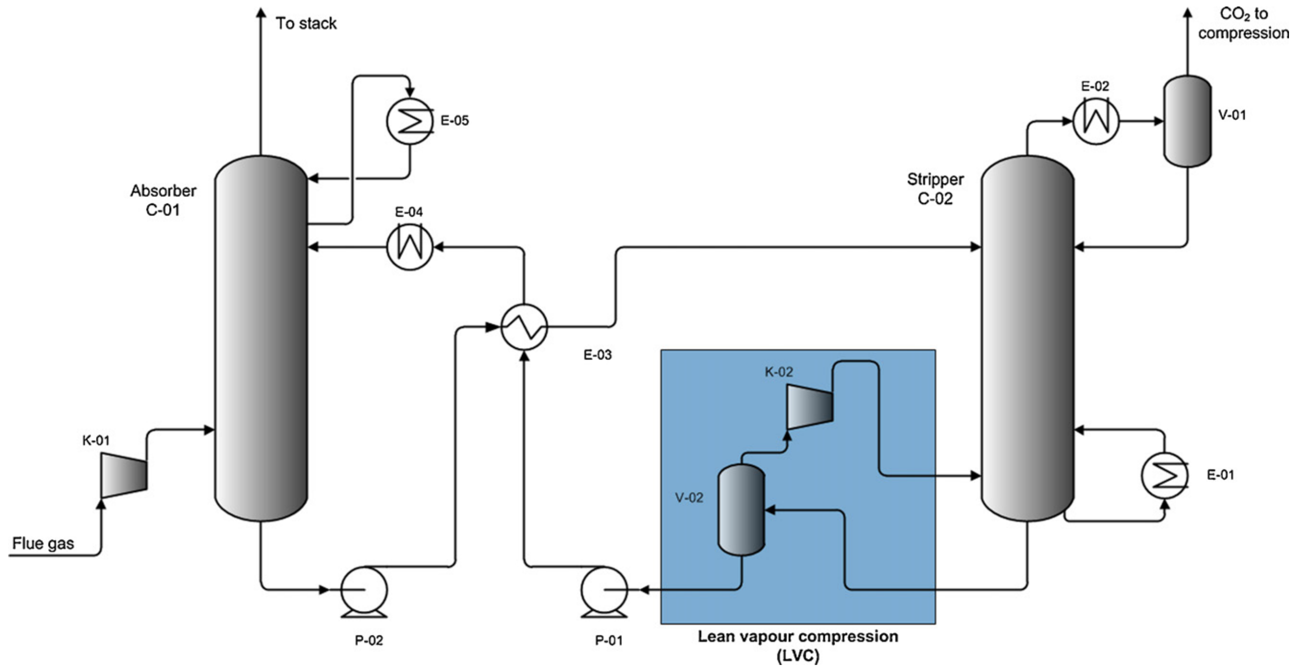

Fig. 1. Standard MEA plant with LVC. Main unit operations in LVC, which are flash vessel (V-02) and compressor (K-02), are highlighted. The rest of the equipment is included in a standard MEA process: flue gas blower (K-01), absorber (C-01), rich pump (P-02), lean pump (P-01), stripper (C-02), reboiler (E-01), condenser (E-03), lean-rich heat exchanger (E-03), cooler (E-04), and condenser drum (V-01).

The addition of the LVC has the following general implications to the process operation:

- The reboiler duty decreases due to the extra stripping vapour coming from the LVC flash vessel.
- Additional electricity is needed to drive the LVC compressor.
- Condenser and LRHX duties decrease due to the loss of heat from the lean stream.
- The stripper needs to accommodate a slightly higher vapour flow compared to that of the standard MEA case.
- Additional equipment is needed, such as a flash vessel, and compressor and a pump.

Two different scenarios can be encountered if LVC is applied to the standard MEA plant: LVC is included in capture plant design or LVC is retrofitted to an existing capture plant.

In the first scenario, it is assumed that the decision of adding an LVC process modification is taken prior to investment. Therefore, the design of the capture plant is affected by the process modification. Design modifications are taken into account to calculate new investment and process operating costs:

- The stripper diameter might be increased to accommodate more vapour flow without losing separation efficiency.
- The heat transfer area might be decreased for both LRHX and condenser when they are designed at a constant temperature approach (pinch).
- The LVC compressor is a turbo-compressor. It has a specific working range. Outside of this range the required duty will be high. The LVC should be operated so, that the efficiency of the compressor remains high.

In the second scenario, it is assumed that the LVC is retrofitted to an existing plant or to an existing designed plant without any modifications in the affected equipment. The LVC compressor can still be designed to operate at its optimal range. However, the heat transfer equipment shall not be modified and remains identical to the standard MEA plant. This will result in relatively higher investment costs for LRHX and condenser, compared with the first scenario.

Both scenarios are investigated in this study. The final pressure in the flash vessel determines the savings in energy and, at the same time, the capital investment. For both scenarios, the reboiler duty and the LVC compressor and pump work have been calculated as a function of flash vessel pressure. Moreover, the extra investment in additional equipment has been estimated as a function of flash vessel pressure. For the first scenario, design modifications to the heat transfer equipment have been also analysed for different flashing pressures in the flash vessel. Nevertheless, the effect on stripper diameter was not considered since the vapour stream generated in the flash vessel is small at the stripper and flashing pressures considered in this study.

2.3. Modelling

Simulations were performed with Aspen Plus® version 7.1. The absorber is modelled with three equilibrium stages and the stripper is modelled with 8 equilibrium stages. This simulation method is an approximation to the phenomena that take place in absorption with chemical reaction. It is based on the minimum number of equilibrium stages, as derived from a pressure – liquid loading plot. The accuracy and applicability of this simulation scheme and its comparison to the rate-based simulation procedure has been already discussed in other papers (Abu-Zahra, 2009). When the reboiler duty is the main parameter of interest, it was found that the simulation results of this approach are in good agreement with the results of the rate-based approach.

The model parameters for both scenarios are listed in Table 1. The CO_2 capture percentage is maintained at 90% capture by manipulation of the solvent flow. The lean and rich solvent CO_2 loadings are a result from the calculations.

CO_2 recovery in the stripper is a design constraint added to the stripper to control the loading of the lean stream leaving this unit operation. It is defined as the ratio (mol basis) of the product

Table 1
List of parameters that are fixed for all simulations of the capture plant.

Parameter	Value
CO_2 capture percentage	90%
Solvent	30 wt% MEA
Absorber	
Feed temperature/pressure (flue gas and solvent)	40 °C/1.1 bar
Equilibrium stages absorption	3
Equilibrium stages washer	2 loops of 3 stages
Pressure drop	90 mbar
LVC compressor	
Compression to	1.9 bar
Total efficiency compressor	77%
Flue gas composition	
CO_2 mol%	13.7
H_2O mol%	12
N_2 + Ar mol%	71
O_2 mol%	3.3
Stripper	
Rich solvent feed stage	3
Equilibrium stages stripping	6
Equilibrium stages wash	2
Pressure of stripper feed	4 bar
Pressure drop	100 mbar
Reboiler pressure	1.8 bar
CO_2 recovery in stripper	0.57
Condenser temperature	40 °C

streams to the feed streams for a specific component (CO_2 in this case). The product stream in the present modelling case is only the lean stream leaving the stripper. Feed streams are the rich stream and the re-compressed vapour entering the stripper.

The energy analysis is performed in equivalent work. The total equivalent work for each simulation is calculated as follows:

$$W_{eq} = \alpha \cdot Q_r + W_p + W_{bc} \tag{1}$$

In the equation above, Q_r is the estimated reboiler duty, W_p is the estimated work for pumps and W_{bc} is the estimated work for absorber blower and LVC compressor. The term α accounts for the loss of turbine power due to steam extraction for the reboiler. This factor depends on the necessary steam quality. Since steam quality remains unchanged in this analysis, a constant value for α of 0.23 has been assumed. This value was derived from power plant and capture plant integration studies by Bolland (2004) for the case where the steam temperature for heating the stripper is in the range of 120–150 °C. For the calculation of pump work pump efficiencies of 85% have been assumed (including mechanical and drivers efficiency). This is in line with similar studies (Franco et al., 2010). For compressor and blower, efficiencies of 77% have been assumed (Bergsma, 2011).

2.4. Economic frame work and analysis

The costs and estimating methodology are directed toward the "study-level" estimate with a nominal accuracy of ±30%. The purpose of the study-level estimate is to compare the different scenarios to the reference case. The estimating procedure relies in the calculation of the mass and energy balances of the different cases (as described in the previous section). The generated information is used in the calculation of main equipment design. Once mass and energy flows and main equipment data are available, equipment cost is estimated.

There are different methods for cost estimation. The most accurate procedure is to request budget quotes from different vendors, specific to the location and conditions of each case. This method is unpractical when the number of cases is high. Instead, historical data from previous projects and/or vendor quotes can be used to estimate equipment cost specific to a given case. This latter option requires the manipulation of cost data to account for differences in scale, construction material, design pressure or time (Peters et al., 2002). Finally, if no historic data points are available, cost data can be also obtained from various models published in the literature (Abu-Zahra et al., 2007; Oexmann and Kather, 2009) or from commercial software packages.

Table 2
Parameters for the economic evaluation of the different cases.

Parameter	Value	Reference
Installation factor	4	Peters et al. (2002)
Interest percentage	8%	Franco et al. (2010)
Turbine power loss to reboiler energy (MWe/MWth)	0.23	Bolland (2004)
Electricity (€/MWh)	50	Bergsma (2011)
Project life time (years)	25	Bergsma (2011)
Compressor life time	10 years	Siemens (2010)
Flash vessel life time	25 years	Peters et al. (2002)
Heat exchangers life time	20 years	Peters et al. (2002)

The approach followed in this work consists in the construction of an in-house model based on a combination of the above mentioned options. At a first stage in model development, several vendors were approached to give quotes for the scale, construction materials and design parameters estimated for the reference case and the optimal case found for each scenario. These quotes were included in a historical database to create significant ground information. At a second stage of model development, different correlations published in the literature and/or suggested by vendors were used to fit the data. As a result of this fitting procedure, a set of price correlations was developed that allows for equipment price estimation under a broad range of operating conditions.

All data in the historical database were manipulated in the following order: material correction, calibration to year 2010 by using the Chemical Engineering Plant Cost Index and currency exchange to Euro. Quotes were received during years 2009 and 2010. The ones from 2009 were calibrated also to the year 2010. All quotes were received in Euro.

The economic analysis is based on calculating the energy savings with respect to the standard MEA plant and calculating the extra investment costs or savings for each particular case. Results are given in net present value (NPV) of the cash flows associated with each case over the entire project life time. The cash flow represents the savings obtained as a function of the flash operating pressure. They are estimated by calculating the difference in utility cost and investment compared to the reference case, according to the formulas:

$$\mathrm{NPV} = \sum_{i=1}^{n} \frac{\mathrm{CF}_i}{(1+\mathrm{IR})^i} \tag{2}$$

$$\mathrm{CF}_i = \Delta\mathrm{Energy}_i - \Delta\mathrm{Depreciation}_i \tag{3}$$

$$\Delta\mathrm{Depreciation}_i = \frac{\Delta\mathrm{TC_{HEX}}}{20} + \frac{\Delta\mathrm{TC_{COMP}}}{10} + \frac{\Delta\mathrm{TC_{VESSEL}}}{25} \tag{4}$$

In the equations above, NPV is the net present value for a given case, CF_i is the cash flow in year i, n is the number of years in the project, IR is the interest rate, $\Delta\mathrm{Energy}_i$ is the difference in utility costs (only steam and electricity) between a given case and the reference case for year i, $\Delta\mathrm{Depreciation}_i$ is the difference of the depreciated total equipment costs between a given case and the reference case for year i and DTCH, DTCC, DTCV, are the difference in total cost of heat exchangers (E-01, E02, E-03), compressor (K-02) and vessel (V-01), between a given case and the reference case respectively.

The parameters used in the calculation of NPV are listed in Table 2. Linear depreciation over the whole equipment life time

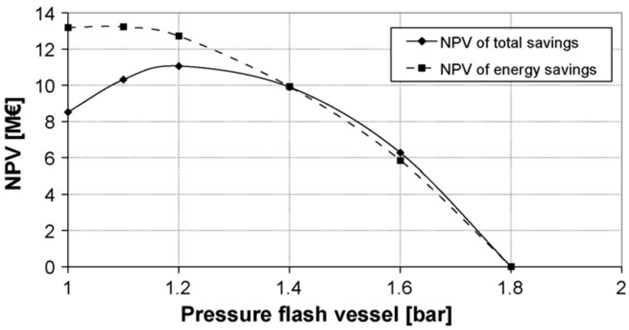

Fig. 2. NPV of total savings, and energy savings against LVC pressure. Case 1: design adapted to LVC operation.

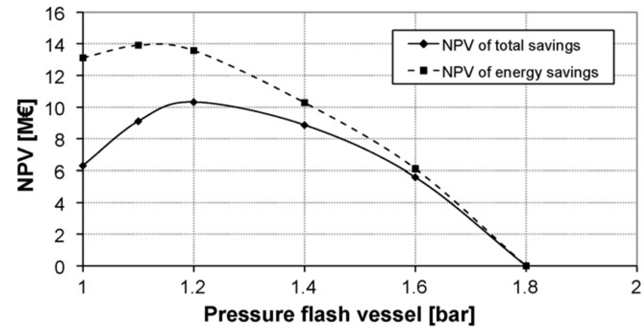

Fig. 3. NPV of total savings, and energy savings against LVC pressure. Case 2: retrofit of LVC to capture plant.

has been used. Selecting the right value for equipment life time that corresponds to the estimated price is crucial in this analysis.

For the LVC compressor the most convenient equipment choice is the Turbo – blower with a pressure ratio up to 2. Most vendors consulted agreed that the life time of the turbo-blower is substantially lower than the other pieces of equipment in this process (Siemens, 2010). Equipment life time might be extended by purchasing a more expensive compressor for which materials are selected to extend the operating life of the equipment. This point is often overlooked in conceptual designs of this type. In this work, the life time of the relevant equipment was adjusted to the type of equipment selected as shown in Eq. (4) and Table 2.

3. Results

3.1. Case 1: capture plant design adapted to LVC operation

In this scenario, the LRHX was modelled at a constant temperature approach (i.e. constant LMTD). Several simulations were conducted with varying LVC flash pressure. Table 3 shows a summary of the equipment duties affected by the LVC.

Due to the loss of latent heat in the lean stream leaving the stripper column, the temperature of the feed stream to this column also decreases. This phenomenon results in a reduction of condenser duty as the flash pressure decreases. The reduction can be as high as 44%. The reboiler duty also decreases with decreasing flash pressures. Energy savings up to 18% can be achieved in the reboiler. However, when the extra pump and compressor work are considered, the energy savings (expressed in total equivalent work) are limited to 7.3%. Moreover, the duty required in the LRHX decreases significantly with decreasing flash pressure. Based on the minimisation of equivalent work, the optimal pressure in the flash vessel is 1.1 bar. However, the changes in equipment duty also have an effect in the cost of equipment.

At each pressure, the investment costs have been estimated using the economic model described in Section 2.4. Table 4 lists the estimated purchased equipment cost of the items considered in this analysis.

Compared to the reference case, costs of the reboiler, LRHX and condenser decrease with decreasing flash pressure. This reduction is explained by the lower duties of reboiler and condenser and the loss of latent heat of the lean hot stream as shown in Table 3. Therefore, the heat transfer area of the mentioned heat exchange equipment can be reduced resulting in lower costs. Obviously, the cost for the LVC compressor, LVC pump and flash vessel increase when the flash pressure decreases due to higher pressure ratio for compression and pumping and higher vapour flow in the flash vessel. However, the cost of LVC equipment is compensated by the lower costs of reboiler, condenser and LRHX at pressures below 1.2 bar. Above that pressure, the cost of LVC equipment becomes dominant.

The net present values (NPV) of the yearly savings due to the application of LVC have been calculated according to the procedure described in Section 2.4. Results are tabulated in Table 5 and depicted in Fig. 2.

Fig. 2 shows the variation of NPV with flash pressure. The dash line represents the hypothetical case, where the investment in the extra equipment is not included in the calculation of annual savings. The solid line represents the actual case, where both investment and energy are included in the calculation of total savings. From the latter case, it can be concluded that the benefit of applying LVC is maximal at a flash pressure of 1.2 bar. Below that pressure the increase in energy savings is small compared to the increase in equipment cost (mainly the compressor).

3.2. Case 2: LVC retrofitted to an existing plant

In this scenario, the LRHX, condenser and cooler were modelled with constant heat transfer area and heat transfer coefficient and equal to the values used in the simulation of the standard MEA plant. Again, several simulations were conducted with varying LVC flash pressures. Table 6 shows a summary of the equipment duties, which are similar to the first scenario (Table 3). The difference in maintaining the heat exchange equipment design unaltered (Table 6) or altered (Table 3) is not really appreciable in the equipment duties. Nevertheless, this study predicts 0.5% lower equivalent work for this scenario around the optimal pressure (1.1 bar).

Regarding the investment costs (Table 7), condenser, reboiler and LRHX remain unchanged and equal to the standard MEA case. However, extra investment cost needs to be added for the LVC compressor and flash vessel, which is also similar to the cost calculated in the previous scenario. Table 8 shows the NPV value for each flash vessel pressure. The NPV results are also plotted in Fig. 3. Energy costs are comparable between both scenarios (Case 1 and Case 2), being the present scenario slightly more favourable. The investment costs are higher in this scenario, resulting in a lower final NPV than in the previous one. Also in this case, the flash pressure that minimizes the equivalent work is slightly lower than the one that maximises the NPV.

4. Discussion

The results of this study show that the application of LVC in both scenarios will reduce the equivalent work necessary for capturing CO_2. Reduction is slightly higher in the case that the LVC is retrofitted to the standard MEA process (7.8% compared to the standard MEA plant) than in the case that the design of the plant is adapted (7.3% compared to the standard MEA plant).

Please cite this article in press as: Sanchez Fernandez, E., et al., Optimisation of lean vapour compression (LVC) as an option for post-combustion CO_2 capture: Net present value maximisation. Int. J. Greenhouse Gas Control (2012), http://dx.doi.org/10.1016/j.ijggc.2012.09.007

Table 3
Overview of equipment duties and equivalent work as a function of LVC flash pressure. Case 1: design adapted to LVC. The total duty of equipment affected by the LVC operation is shown. Equipment names are followed by the corresponding tag in Fig. 1.

Flash pressure (bar)	Reboiler (E01) (MWth)	Pump (P01) (MWe)	Compressor (K02) (MWe)	LRHX (E03) (MWth)	Condenser (E02) (MWth)	Equivalent work (MJ/t CO_2)
1.8	172	0.000	0.0	197	60	819.4
1.6	164	0.182	0.2	166	48	789.2
1.4	158	0.197	0.8	153	42	773.3
1.2	150	0.214	2.0	149	40	760.4
1.1	146	0.222	2.8	142	37	758.1
1.0	141	0.227	4.0	132	34	759.2

Table 4
Affected purchased equipment cost as a function of LVC flash pressure. Equipment names are followed by the corresponding tag in Fig. 1. Only significant changes in equipment cost are considered. Note that the numbers given in this table do not include installation cost. Installation costs are included when calculating depreciation.

Flash pressure (bar)	Reboiler (E01) (M€)	Flash vessel (V02) (M€)	Compressor (K02) (M€)	LRHX (E03) (M€)	Condenser (E02 and V01) (M€)	$\Delta(TC_i)$[a] (M€)
1.8	1.30	0.00	0.0	1.11	0.31	0.00
1.6	1.25	0.11	0.1	0.79	0.25	−0.25
1.4	1.20	0.20	0.3	0.65	0.22	−0.20
1.2	1.14	0.32	0.6	0.61	0.21	0.20
1.1	1.11	0.39	1.0	0.54	0.2	0.47
1.0	1.07	0.47	1.4	0.44	0.18	0.86

[a] Total change in equipment cost: difference in total equipment cost between the reference case (at 1.8 bar), which is 2.72 M€ and any other given case.

Table 5
Calculation of NPV as a function of flash pressure. Case1: design adapted to LVC. Yearly energy cost, energy savings, equipment depreciation difference and NPV for each operating flash pressure.

Flash pressure (bar)	Energy cost (M€/y)	Energy savings[a] (M€/y)	Depreciation difference[b] (M€/y)	NPV of energy savings[c] (M€)	NPV of total savings[d] (M€)
1.8	16.77	0.00	0.00	0.0	0.0
1.6	16.23	0.55	−0.04	5.9	6.3
1.4	15.84	0.93	0.00	9.9	9.9
1.2	15.58	1.19	0.15	12.7	11.1
1.1	15.54	1.24	0.27	13.2	10.3
1.0	15.54	1.24	0.44	13.2	8.5

[a] Difference in energy cost between the reference case (at 1.8 bar) and any other given case.
[b] Equipment depreciation difference calculated with Eq. (4).
[c] NPV calculated with Eq. (2) without taking into account the depreciation difference.
[d] NPV calculated with Eq. (2) taking into account energy savings and depreciation difference.

Table 6
Overview of equipment duties and equivalent work as a function of LVC flash pressure. Case 2: retrofit scenario. The total duty of equipment affected by the LVC operation is shown. Equipment names are followed by the corresponding tag in Fig. 1.

Flash pressure (bar)	Reboiler (E01) (MWth)	Pump (P01) (MWe)	Compressor (K02) (MWe)	LRHX (E03) (MWth)	Condenser (E02) (MWth)	Equivalent work (MJ/t CO_2)
1.8	172	0.000	0.0	197	60	817.5
1.6	164	0.183	0.2	182	53	789.6
1.4	157	0.195	0.8	168	47	770.6
1.2	149	0.214	1.9	151	41	755.5
1.1	145	0.223	2.8	144	38	754.0
1.0	141	0.228	3.9	144	34	757.6

Table 7
Purchased equipment cost as a function of LVC flash pressure. Equipment names are followed by the corresponding tag in Fig. 1. Only significant changes in equipment cost are considered. Note that the numbers given in this table do not include installation cost. Installation costs are included when calculating depreciation.

Flash pressure (bar)	Reboiler (E01) (M€)	Flash vessel (V02) (M€)	Compressor (K02) (M€)	LRHX (E03) (M€)	Condenser (E02 and V01) (M€)	$\Delta(TC_i)$[a] (M€)
1.8	1.30	0.00	0.00	1.11	0.31	0.00
1.6	1.30	0.12	0.08	1.11	0.31	0.19
1.4	1.30	0.21	0.25	1.11	0.31	0.45
1.2	1.30	0.32	0.63	1.11	0.31	0.95
1.1	1.30	0.39	0.97	1.11	0.31	1.35
1.0	1.30	0.46	1.41	1.11	0.31	1.87

[a] Total change in equipment cost: difference in total equipment cost between the reference case (at 1.8 bar), which is 2.72 M€ and any other given case.

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Table 8
Calculation of NPV as a function of flash pressure. Case 2: retrofit scenario. Yearly energy cost, energy savings, equipment depreciation difference and NPV for each operating flash pressure.

Flash pressure (bar)	Energy cost (M€/y)	Energy savings[a] (M€/y)	Depreciation difference[b] (M€/y)	NPV of energy savings[c] (M€)	NPV of total savings[d] (M€)
1.8	16.77	0.00	0.00	0.0	0.0
1.6	16.20	0.57	0.05	6.1	5.6
1.4	15.81	0.96	0.13	10.3	8.9
1.2	15.50	1.27	0.30	13.6	10.3
1.1	15.47	1.30	0.45	13.9	9.1
1.0	15.54	1.23	0.64	13.1	6.3

[a] Difference in energy cost between the reference case (at 1.8 bar) and any other given case.
[b] Equipment depreciation difference calculated with Eq. (4).
[c] NPV calculated with Eq. (2) without taking into account the depreciation difference.
[d] NPV calculated with Eq. (2) taking into account energy savings and depreciation difference.

Table 9
Energy savings of LVC as reported in other studies.

Reference	Energy savings[a] (%)	ΔP^b (bar)	Stripper pressure (bar)
Karimi et al. (2011)	8.39–9.36	0.85–0.9	2
Le Moullec and Kanniche (2011)	7.78	1.25	2.5
This work	7.3–7.8	0.6	1.8

[a] Reduction in equivalent work taking a standard MEA plant as basis.
[b] Difference between stripper pressure and LVC flash pressure.

Table 9 shows the comparison of these results to similar studies. The energy savings are expressed in difference in equivalent work between the reference case (generally similar to the standard MEA plant described in this work) and the LVC configuration. The results of this study are in line with the finding of others. The difference in numbers is related to the different conditions adopted for the stripper.

The operating flash pressure that minimizes equivalent work is 1.1 bar for both scenarios. If the reduction in energy was the only considered criterion for optimisation, the best scenario would be the retrofit of LVC to the standard MEA plant, because the reduction in energy is more significant in this scenario. However, when the trade-off between energy savings and additional capital investment is taken into account, it is found that the optimal flash pressure is 1.2 bar, which implies a pressure difference with the stripper of 0.6 bar. Moreover, when the NPVs of total savings for both analysed scenarios are compared, the best result is achieved when the plant design is adapted to the LVC addition. This example shows the efficacy of this approach.

The differences between the two scenarios are small because the stripper pressure was fixed to 1.8 bar. Therefore, there is small range of pressures to be considered in the optimisation. Nevertheless, higher differences can be expected when the difference between the operating pressure in the stripper and the flash vessel is increased. In that particular case, the approach followed in this work will lead to a better optimisation of operating conditions rather than the approach of minimising equivalent work.

Moreover, the turbine power loss ratio is a very important factor in this type of analysis. For a very well integrated power plant the turbine power loss ratio could be as low as 0.20 (MWe/MWth). This means that the energy savings will be contributing less to the NPV calculated, while the cost of the compressor will remain identical. This fact contributes to the disparity of the pressure that minimises equivalent work and the one that maximises the NPV. In the case of a poorly integrated power plant, this factor could be as high as 0.24 (MWe/MWth). The energy becomes a more dominant factor contributing greatly to the NPV. For this situation, the pressure that minimises the equivalent work will approach the pressure that minimises the NPV.

Besides the equivalent work and NPV minimisation, other aspects should be also addressed to better understand the LVC benefits. One of these aspects is operability and possible LVC trip (due to failure or maintenance).

When the LVC is turned off, the operation of reboiler, lean-rich heat exchanger and condenser is affected. The resulting plant performance depends on whether the plant design was adapted to LVC operation (case described in Section 3.1) or the LVC was just retrofitted to an existing plant (case described in Section 3.2). The following analyses both scenarios.

In the case that plant design was adapted for LVC, the mean temperature difference in the LRHX will rise upon LVC shut down due to the higher energy in the lean stream. The reboiler duty also increases. There will be more flow to the condenser, which will increase its pressure drop, having an effect in the compression train. When not counteracted by adjustment of the reboiler and condenser, shut-down of the LVC will lower the capture percentage of the plant.

In the case that the LVC was just retrofitted to the plant, operation returns to nominal values when the LVC is shut off. The plant can be operated normally by increasing the steam supply to the reboiler and the cooling duty of the condenser and lean liquid cooler. This makes a capture plant with retrofitted LVC more flexible than the option where the plant is adapted to the lower duties and gives and additional advantage to this option.

5. Conclusions

In this work, the application of lean vapour compression to a standard MEA carbon capture plant was investigated. The worked focussed on the financial analysis of this process modification and it can be used to make an informed decision on whether this process modification should be added or not to a current situation.

Two scenarios were analysed. In the first scenario, the design of the capture plant was modified to accommodate the changes in duties resulting from LVC application. In the second scenario, the design was unchanged and the affected equipment was not modified.

Under the conditions investigated, it was found that LVC will result in savings for both scenarios chosen for implementation. The operating conditions were optimised based on the NPV of the process design. It can also be concluded that the right choice of operating pressure in the LVC maximises the NPV. Under the investigated conditions, the optimal operating pressure is 1.2 bar for both scenarios. It was also found, that the optimisation of operating conditions based only on energy analysis could lead to a different result.

Finally, the most attractive scenario from an economic point of view is the one where the capture plant is fully adapted. However, the retrofit scenario has other advantages, not quantified in this financial analysis, such as the flexible shut down of the LVC (for maintenance or reparation). This can make this scenario more attractive from a reliability point of view.

Acknowledgments

The authors would like to thank the CATO-2 program, coordinators and sponsors for financing this research.

References

Abu-Zahra, 2009. Carbon dioxide capture from flue gas: development and evaluation of existing and novel process concepts. PhD thesis. Delft University of Technology. Available from http://repository.tudelft.nl/

Abu-Zahra, M.R.M., Niederer, J.P.M., Feron, P.H.M., Versteeg, G.F., 2007. CO_2 capture from power plants. Part II. A parametric study of the economical performance based on mono-ethanolamine. International Journal of Greenhouse Gas Control 1, 135–142.

Bergsma, E., 2011. CATO-2 Deliverable WP 1. 1A – D01D. Public summary will be available at: http://www.co2-cato.org/publications

Bolland, O., 2004. Common framework of the modelling basis. ENCAP report D6.1.1/D4.1.1.

Davison, J., Thambimuthu, K., 2009. An overview of technologies and costs of carbon dioxide capture in power generation. Proceedings of the Institution of Mechanical Engineers, Part A: Journal of Power and Energy 223 (3), 201–212.

Finkenrath, M., 2011. Cost and performance of carbon dioxide capture from power generation. Working paper. International Energy Agency.

Franco, F., Anantharaman, R., Bolland, O., Booth, N., van Dorst, E., Ekstrom, C., Sanchez, E., Macchi, E., Manzolini, G., Prins, M., Pfeffer, A., Rezvani, S., Robinson, L., Abu-Zahra, M.R.M., 2010. Common framework and test cases for transparent and comparable techno-economic evaluations of CO_2 capture technologies – the work of the European Benchmarking Task Force. In: Proceedings of GHGT-10, Amsterdam (NL), September 2010.

Gibbins, J., Chalmers, H., 2011. Techno-economic assessment of CO_2 capture retrofit to existing power plants. Energy Procedia 4, 1835–1842.

Huizeling, E., Weijde, G., 2011. Non-confidential FEED study report. Special report for the Global Carbon Capture and Storage Institute. Download from: www.road2020.nl

Karimi, M., Hillestad, M., Svendsen, H., 2011. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO_2 capture. Chemical Engineering Research and Design 89 (8), 1229–1236.

Le Moullec, Y., Kanniche, M., 2010. Optimisation of flow diagram modifications and their interaction for an efficient MEA based CO_2 post combustion capture. Chemical Engineering Transactions 21, 175–180.

Le Moullec, Y., Kanniche, M., 2011. Screening of flowsheet modifications for an efficient monoethanolamine (MEA) based post-combustion CO_2 capture. International Journal of Greenhouse Gas Control 5 (4), 727–740.

Oexmann, J., Kather, A., 2009. Post-combustion CO_2 capture in coal-fired power plants: comparison of integrated chemical absorption processes with piperazine promoted potassium carbonate and MEA. Energy Procedia 1 (1), 799–806.

Peters, M., Timmerhaus, K., West, R., 2002. Plant Design and Economics for Chemical Engineers, 5th ed. Mc Graw Hill, New York, ISBN 978-0072392661.

Rubin, E.S., Chen, C., Rao, A.B., 2007. Cost and performance of fossil fuel power plants with CO_2 capture and storage. Energy Policy 35 (9), 4444–4454.

Siemens. Personal communication with M. Rost in 2010.

Van Wagener, D.H., Rochelle, G.T., 2011. Stripper configurations for CO_2 capture by aqueous monoethanolamine. Chemical Engineering Research and Design 89 (9), 1639–1646.

Wang, M., Lawal, A., Stephenson, P., Sidders, J., Ramshaw, C., 2011. Post-combustion CO_2 capture with chemical absorption: a state-of-the-art review. Chemical Engineering Research and Design 89 (9), 1609–1624.

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