Simulation-based Techno-economic Evaluation for Optimal Design of CO_2 Transport Pipeline Network

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Abstract

For large volumes of carbon dioxide (CO_2) onshore and offshore transportation, pipeline is considered the preferred method. This paper presents a study of the pipeline network planned in the Humber region of the UK. Steady state process simulation models of the CO_2 transport pipeline network were developed using Aspen HYSYS®. The simulation models were integrated with Aspen Process Economic Analyser® (APEA). In this study, techno-economic evaluations for different options were conducted for the CO_2 compression train and the trunk pipelines respectively. The evaluation results were compared with other published cost models. Optimal options of compression train and trunk pipelines were applied in an optimal case. The overall cost of CO_2 transport pipeline network was analyzed and compared between the base case and the optimal case. The results show the optimal case has an annual saving of 22.7 M€. For the optimal case, levelized energy and utilities cost is 7.62 €/t-CO_2, levelized capital cost of trunk pipeline is about 8.11 €/t-CO_2 and levelized capital cost of collecting system is 2.62 €/t- CO_2. The overall levelized cost of the optimal case was also compared to the result of another project to gain more insights for CO_2 pipeline network design.

Keywords: CCS, CO_2 transport, Pipeline transport, Process simulation, Economic evaluation

1. Introduction

1.1. Background

Reducing CO_2 emissions to achieve greenhouse gas emissions target is a significant challenge both technically and commercially. Power generation from fossil fuel (e.g. coal and natural gas) fired power plants is the single largest source of CO_2 emissions [1-2]. Fossil energy is projected to remain a major source of energy in the near future. Therefore, CCS will make a vital contribution towards reducing CO_2 emissions from power plants and meeting the global CO_2 emission target [3].

CCS is a process of capturing CO_2 from large industrial sources and transporting it to a storage site, to mitigate CO_2 emission to the atmosphere. In the transport section of CCS, CO_2 is compressed and transported from capture plants to storage sites by pipeline, ship or tanker trucks mainly depending on the distance. Pipelines are the preferred method for onshore and offshore transport of large volumes of CO_2 [4-5]. The CO_2 pipeline transport for enhanced oil recovery (EOR) is a mature technology. Several millions of tonnes of CO_2 are already transported for EOR purpose by pipelines in the USA and Canada. In 2050, to achieve the carbon emission target, the global CO_2 captured is about 7Gt/y [6]. This is a much larger amount than about 50 Mt/y transported in pipelines for EOR in the USA [7]. A massive investment and a significant effort would be required to build a CO_2 infrastructure of such scale.

There are many CCS demonstrations projects being planned or implemented over the world. In the UK, the Humber region offers the good opportunities for CCS deployment as it is not only the biggest CO_2 emission area in the UK, but also the area with easy reach to the CO_2 offshore storage sites in the North Sea [8]. In the Humber region, UK, the Don Valley Power Project and the White Rose CCS demonstration project were approved in 2009 and 2013 respectively. These projects are expected to provide the basis for the development of a pipeline network to support a cluster of CCS plants in this area.

1.2. Previous studies

Pipelines have been used to transport CO_2 in gaseous and dense (including subcooled liquid or supercritical) phases. The dense phase is regarded as the most energy-efficient condition due to its high density and low viscosity [9-10]. Consequently, current operating practice for CO_2 pipelines is to maintain the pressure well above the critical pressure. The impurities in CO_2 stream have great impacts on the design, operation and optimisation studies [11-13]. Seevam et al. [14] studied the impact of the impurities on phase behaviour and density of CO_2. The presence of the impurities may result in the formation of gaseous CO_2 or two-phase CO_2 flow inside the pipelines. The water content in the CO_2 stream results in the risk of hydrate, which will pose a complex problem involving phase transient and pipeline blockage [13, 15-16]. Therefore, before the transport, the CO_2 stream has to be conditioned to remove impurities such as water vapour, H_2S, N_2, methane, O_2, hydrocarbons and free water [17-18].

Steady state simulations and analysis were carried to calculate pressure drop, temperature profile and mass flow in the pipelines. Zhang et al. [9] studied the density and pressure profiles of CO_2 stream along the length of the pipeline with different inlet temperatures. Maximum safe pipeline length was determined for different inlet temperatures. In the study of Nimtz et al. [19], the model includes the pipeline and an injection well for pure CO_2 stream. The profiles of pressure, temperature, density and flow velocity were presented for several cases and the phase change was found and discussed. Regarding the dynamics of pipeline systems, there is little work reported in the literature. Liljemark et al. [20] developed a pipeline transfer function model to evaluate phase transition of the transported CO_2 mixture. Operation scenarios of pipeline cooling, load change, start-up, shut-down and compressor trip were simulated. Chaczykowski and Osiadacz [21] built a first principle single-phase compressible flow model, suitable for supercritical and dense-phase calculations, to examine the hydraulic parameters of the CO_2 pipeline. However, these simulations were performed for a single CO_2 emission source without intermediate boosters. This may not reflect realistic operating scenarios for a typical CO_2 pipeline network system.

The cost of the CO_2 pipeline transport can become significant when the distances between the storage locations and the emission sources are more than a few hundred kilometres. Collecting CO_2 mixture from several emitters into trunk pipelines is more cost-effective than the use of separate pipelines [5, 22]. As a part of economic evaluation of CCS deployment, some research efforts were given on the cost estimate of CO_2 pipeline transport. Van den Broek et al. [23], Heddle et al. [24] and Element-Energy [25] used a linear cost related with diameter and length of the pipelines to calculate investment costs. Gao et al. [26] developed a cost model based on the weight of the pipeline, which is specific for the Chinese market. In the report of IEA-GHG [27], six different kinds of coefficients, for 600#, 900# and 1500# ASNI class and onshore/offshore pipelines, were used for the operating and maintenance costs calculation of CO_2 transport pipelines. McCoy and Rubin [10] developed a cost equation for pipeline transport with different parameters for each cost category (material, labour, right of way and miscellaneous costs) for different regions of the USA. Dahowski et al. [28]and McCollum and Ogden [29] built their linear cost equations only based on the flow rate of CO_2 stream and the length of the pipelines.

The cost of transporting (not include the cost of compression) the CO_2 by a 100 km onshore pipeline was estimated by the Global CCS Institute (GCCSI) at 0.46–1.55 €/t CO_2 [30]. However, large ranges for capital and levelized costs of CO_2 transportation were found for a given diameter [29, 31-33]. For example, Knoope et al. [33] came up with a cost range of 0.6–11 M€/km for a 0.91 m-diameter pipeline after comparing seven different models. The reason for the prediction of large cost ranges may be down to the fact that most estimation models are empirical correlations developed on the basis of historical cost data of natural gas pipeline projects and do not reflect the process conditions of different CO_2 pipeline projects. Furthermore, most of the cost models in literature focus only on pipeline costs, but an integral assessment of CO_2 transport costs should go beyond pipelines by including entry compressors and boosters stations. For the optimal design of CO_2 pipeline networks, it is necessary to have techno-economic evaluation models linking compression, pipelines and booster station to each other.

1.3. Aim of the paper and novel contribution

This paper is aimed to carry out techno-economic evaluation based on steady state simulation to explore the optimal design of the CO_2 transportation pipeline network, in order to reduce the cost of CCS deployment. Servicing this aim, the objectives of this study are: 1) to develop a detailed process model including the CO_2 compression trains, collecting pipelines, onshore trunk pipeline, booster pump station and offshore trunk pipeline. The CO_2 capture plant, gas conditioning plant, injection and storage are not included in this study; (2) to carry out simulation-based techno-economic evaluations for different design options of CO_2 compression train and trunk pipelines respectively and then to select the optimal options. The results will be compared with other published cost models; (3) to compare the overall costs of CO_2 transport pipeline network of the base case and the optimal case. The cost results will be compared with the published study on another project.

There are two novelties claimed in this study: (1) the process model in this study is more detailed than other published models of the CO_2 transport pipeline network. The model in this study includes CO_2-rich streams coming from two emitters, CO_2 compression trains, the collecting pipelines, the onshore/offshore trunk pipelines and the booster pump station; (2) the techno-economic evaluations are based on detailed process simulations which links the compression, booster pump station and collecting and trunk pipelines together. This is in contrast with many published studies in the same area because those studies were based on correlations or historical data. This evaluation method offers more accurate cost estimations for decision making support for selecting detailed technical options.

2. Pipeline network system

In the Humber region of the UK, there are two advanced proposals for power station with CCS developments that utilise the trunk pipelines: The Don Valley Power Project (DVPP), promoted by 2Co Energy; and the White Rose CCS Project, promoted by Capture Power Limited. DVPP will use pre-combustion carbon capture technology (physical absorption with selexol solvent) at a new-build integrated gasification combined cycle (IGCC) power plant of 920 MWe gross output [34]. The White Rose CCS project is a demonstration project of an oxyfuel power plant of 450 MWe gross output [35]. National Grid will construct and operate the CO_2 transport pipelines and the permanent CO_2 undersea storage facilities at a North Sea site [36]. One statement is that, except the special references, the input information of the pipeline network in this study (including the route and parameters of the pipelines, the CO_2 mixture properties and specifications and the operating constrains) are provided/agreed by National Grid for this study.

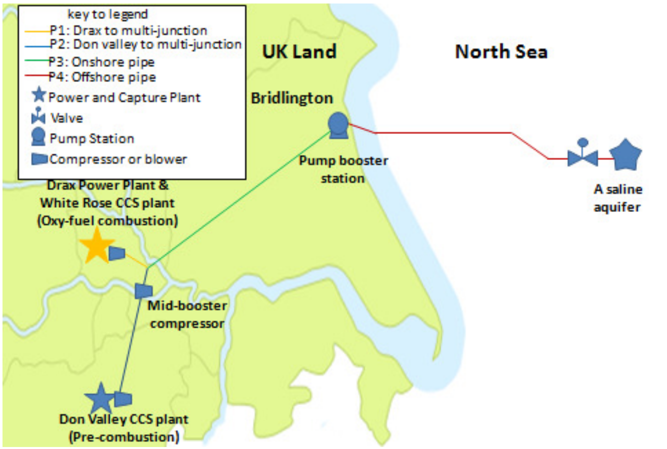

Figure 1. The pipeline sketch for the Humber case study

Figure 1 shows the proposed route corridor of the pipeline network in this study. The CO_2 captured from DVPP will be transported in gaseous phase at a maximum allowable operating pressure (MAOP) of 35 bar, and would then be boosted to dense phase by a compressor near the multi-junction site, before joining the dense phase CO_2-rich stream from the White Rose CCS plant. The combined CO_2-rich stream will then be transported in dense phase via an onshore trunk pipeline with a MAOP of 136 bar. The onshore pipelines are buried under ground 1.2 m. A booster pumping station located near the coast will boost the pressure of the CO_2-rich stream before it is transported in the offshore trunk pipeline with a MAOP of 186 bar to a saline aquifer storage site more than 1 km beneath the bed of the North Sea. Table 1 presents the key parameters of the pipelines. The material of pipelines is carbon steel and the size of pipelines is following ANSI standard.

Table 1. Parameters of the pipelines

Emitter	Flow rate range	Collecting pipelines		Onshore trunk pipeline		Offshore trunk pipeline	
		Length	Nominal Diameter	Length	Nominal Diameter	Length	Nominal Diameter
	Mtpa	km	inch	km	inch	km	inch
Don Valley	0.91 to 6.27	15	30	71	24	91	24
White Rose	0.61 to 2.65	5	12				

An entry specification for the CO_2-rich stream is needed to define the acceptable range of composition, taking into account safety, impact on pipeline integrity and hydraulic efficiency [13]. For this design of the pipeline network, the entry specification agreed between the power plants operators and National Grid was defined to be 96 mole% CO_2 and a mixture of nitrogen, oxygen, hydrogen, argon and methane with hydrogen limited to 2.0 mole% and oxygen limited to 10 ppmv. The entry temperature is not higher than 20 ^{o}C.

3. Process simulation model development and economics evaluation methodology

For a real CO_2 pipeline network project described in section 2, the techno-economic evaluation should be more detailed and realistic. Therefore, insights can be obtained for optimal design of the pipeline network. In this study, process simulation models were developed in Aspen HYSYS®. Then the simulation results were exported into APEA for economic evaluation. This techno-economic evaluation tool offers a more rigorous capital, O&M and energy cost assessment than other cost estimation models.

3.1. *Process simulation model development for the base case*

3.1.1. *Physical property method*

The cubic equation of state (EOS) such as Soave-Redlich-Kwong (SRK) [37] and Peng-Robinson (PR) [38] has been widely used to calculate the physical properties of the CO_2 and impurities [39]. More complex EOS such as Lee Kesler [40], the Statistical Associating Fluid Theory (SAFT) [41-42], Span and Wagner (SW) [43] and GERG [44] were used in recent studies . SW is accurate for pure CO_2 as it was specially developed for pure CO_2. But it is difficult to generalize for multi-component mixture [45] because it contains many terms, some of which are too complex exponential for computation [46]. Molecular-based SAFT is an attractive EOS for CO_2 including impurities because of better performance than other models for predicting thermodynamic properties of several complex mixtures. SAFT-VR, one of modifications of original SAFT, is used for CO_2 capture process [47-48]. But SAFT is not yet used in the published literatures focusing on the dense phase pipeline transport of the CO_2 and impurities. GERG is the international reference equation of state for natural gas. The accuracy of GERG EOS claims to be very high covering a large part of the T/P range for CCS application. GERG was used in recent studies emphasizing on the transient behaviors of the CO_2 and impurities in dense phase pipeline transport [20-21]. However the average absolute deviations (AAD) of the liquid volume of CO_2 mixtures could reach up to 18%, reported by Li et al. [49].

There is no consensus in the literature regarding the best EOS for the design of the CO_2 pipeline transport. PR EOS was chosen in some studies [9, 14, 50] giving reasonable results for properties of the CO_2 and impurities. Li et al. [39, 51] concluded calibrated binary interaction parameters based on the experimental data improve the accuracy of EOS after comparing the results generated with the literature k_{ij} and the new calibrated k_{ij}. His later study [49] presents SAFT shows a better accuracy than PR for volume calculation, but PR shows an advantage in VLE calculations. Diamantonis et al. [45] compared the results of several EOS with experimental data and found that PR EOS is of reasonable accuracy, even when compared with more advanced EOS such as SAFT and PC-SAFT, when calibrated binary interaction parameters are used.

In this study, PR EOS with calibrated binary interaction parameters has been used considering both the accuracy and the simplicity. Table 2 lists the calibrated binary interaction parameters for PR-EOS used in this study. The AADs between the calculations of PR EOS and the experimental data were listed for corresponding k_{ij} values. For calibration of k_{ij} of CO_2-H_2, there is no good agreement among the available experimental data and there is no liquid volume experimental data [49].

Table 2. AAD% between experimental data and PR-EOS for corresponding k_{ij} values

	k_{ij}	Bubble pressure			Liquid volume			Reference
		Temperature (K)	Pressure (Mpa)	AAD%	Temperature (K)	Pressure (Mpa)	AAD%	
CO_2 - N_2	-0.007	220-301	1.4-16.7	3.73	209-320	1.4-16.7	1.54	[39, 45]
CO_2 - Ar	0.141	288	7.5-9.8	2.32	288	2.4-14.5	1.83	[45]
CO_2 – H_2	0.1470	290.2	5.0-20.0	5.6%	-	-	-	[52]

One weakness of PR EOS reported by E.ON's report [53] is that it is very accurate in the near-critical region. This study focuses on the techno-economic evaluations based on steady state simulations. For the trunk pipeline transport section, the CO_2-rich stream is in the subcooled liquid phase. The temperature range is from 4^{o}C to 20^{o}C and the pressure range is from 101 bar to 150 bar, which is far away from the critical region of the CO_2. In this T/P range, the deviation of pure CO_2 density is from -4.8% to 0.1% for the calculations of PR compared to the calculations of SW according to the comparison results in E.ON's report.

3.1.2. *Assumptions, constraints and inputs*

The maximum entry flow rates from both the White Rose plant and the Don Valley plant and the highest ambient temperature (ground temperature is14 ^{o}C and sea water temperature is 16 ^{o}C) were chosen as the base case. This is considered as the worst case scenario with respect to the energy requirement since it would require the highest entry pressure and the greatest boosting pressure at the pump station. Figure 2 shows the flowsheet of pipeline network developed in this study.

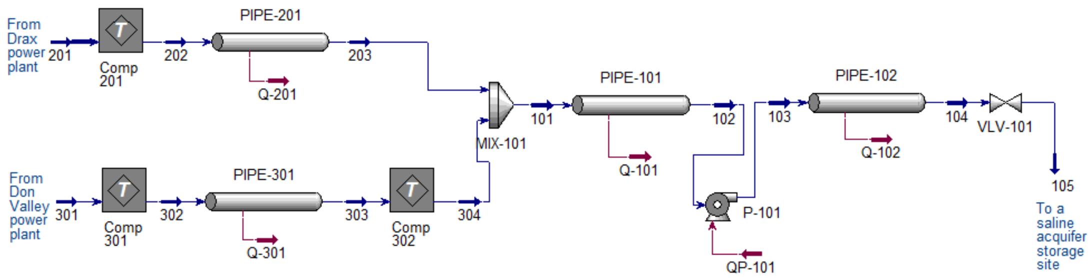

Figure 2. The flowsheet of pipeline network in Aspen HYSYS®

The assumptions made for the pipeline network model are as follows: 1) the pressure drops across valves and other fittings are negligible. 2) the adiabatic efficiencies of compressors and pumps used in this model are fixed at 75%.

The above two assumptions can be justified as follow: 1) the numbers and details of the valves and fittings of the pipelines are unknown at this stage of the project. On the other hand, the pressure drop of the valves and fittings is much smaller than the pressure drop of pipelines itself at such long length; 2) 75% adiabatic efficiency is also used for compressors and pumps in the study of McCollum & Ogden [29]. The efficiency of a compression and pump is the ratio of the useful work done by the equipment on the fluid to the work put into the equipment by the motor. So the efficiency does not affect the hydraulic performance of the pipelines and only linearly affects the actual energy requirement of the compressors and pumps.

The pressure settings of key sections are based on two operational constraints. Firstly, the entry pressure (the outlet pressure of the compressor at each capture plant) should be high enough to maintain a minimum pipeline operating pressure of 101barg to avoid two phase flow in the trunk pipeline. The pressure is obviously higher than the critical pressure of the CO_2 as it considers the impacts of elevation changes of the pipelines and the impurities of the CO_2 mixture. Secondly, a constant injection pressure of 126 bar is specified to satisfy the injection rate. In reality, the required injection pressure at the offshore storage site will rise over the lifetime of the operation with injection pressures below 126 bar being sufficient in the initial phase.

The input and boundary conditions for the base case are specified in Table 3. The configuration of the compressors including the number of stages and the exit temperature of the intercoolers refers to the study of Zhang et al. [9].

Table 3. Input and boundary conditions of the base case

	unit	the White Rose plant	the Don Valley plant
Capture type	-	oxy-fuel	pre-combustion
Composition of CO_2 -rich stream	mol%	96%CO_2, 2%N_2, 2%Ar	96%CO_2, 2%N_2, 2%H_2
Flow rate[a]	t/h	334.596	791.667
Suction pressure of compression	bar	1	1
Suction temp. of compression	°C	20	20
Number of compression stages	-	5	4
Exit pressure of compression	bar	136.0	35
Exit temp. of compression	°C	20	20
Number of mid-compressor stages	-	-	2
Exit pressure of mid-compression	bar	-	136.0
Trunk pipelines entry temperature	°C	20	20.0
Differential pressure of pump station	bar	26.55	
Offshore platform arrival pressure	bar	126.0	

a. *The flow rate of the CO_2-rich stream from each power plant is calculated based on the annual maximum CO_2 loading in Table 1, assuming 330 working days per year.*

3.1.3. Comparing the results of the Aspen HYSYS® model to the PIPEFLO® model

For large scale CO_2 pipeline network simulations, operational or experimental data for model validation purpose is not available as the projects considered are currently only in the planning stage. In this case study, the simulation results using Aspen HYSYS® were compared with the simulation results using another software package, PIPE-FLO®, in order to make a brief validation. GERG EOS was used in PIPE-FLO® for the validation simulations. Table 4 shows the comparison results of Aspen HYSYS® and PIPE-FLO®, which appear to be in good agreement.

Table 4. Comparison of the simulation results

	Entry pressure at White Rose	Entry pressure at Don Valley	DP of mid-booster for Don Valley	DP of pump station	Arrival pressure
	bar	bar	bar	bar	bar
Aspen HYSYS®	120.5	35	86.92	43	126
PIPEFLO®	120.2	35	86.70	42.4	126
Relative difference	0.25%	-	0.25%	1.40%	-

3.2. Economic evaluation methodology

Economic evaluations are conducted using APEA V8.0 using data from the 1st quarter of 2012. APEA becomes an industry-standard tool known to be far more accurate than correlation-based economic approaches [54] and is used for engineering design of many projects. APEA includes design procedures and costs data for hundreds of types of materials of projects. A bottom-up approach is used in APEA. When the simulation models are exported into APEA, the unit operations are mapped to separate equipment cost models, which are then sized and designed according to relevant design codes.

The total cost includes capital expenditure (CAPEX) and operational expenditure (OPEX). OPEX can be split into fixed OPEX (operating and maintenance (O&M) cost) and variable OPEX (mainly the energy and utilities cost) [27]. In this study, for a clearer comparison, the annual cost and the levelized cost (per tonne of CO_2) were used. The total annual cost was split into annualized capital investment cost (capital return factor is 0.15 [29]), annual O&M cost and annual energy and utilities cost. In consistency with that, the levelized cost was split into levelized capital cost, levelized O&M cost and levelized energy and utilities cost.

To harmonize results for comparison with other studies, the following assumptions are made: 1) the project begins in January 2012; 2) all costs are corrected to €2012 using the average inflation index of worldwide; 3) the captured CO_2 mixture has neither economic value nor disposal cost; 4) cooling water is sourced from a nearby body of water at the cost of pumping and operation of a cooling tower. Other important cost inputs are provided in Table 5, with the costs given in Euro. The utility unit prices in Table 4 are from the database of 1st quarter of 2012 of APEA.

Table 5. Key economic evaluation cost inputs

Description	Unit	
Electricity price	€/kW	0.0775
Cooling water price	€/m^3	0.0317
The price of refrigerant-Freon 12	€/t	0.17
Carbon steel price	€/kg	500
Interest rate	%	15
Contingency	%	5
Project economic life	year	30

3.3. Work flow of techno-economic evaluation

For a given base case, the simulation model of whole pipeline network was developed first to check its accuracy and to gain basic data of streams and processes. The evaluations for different technical options go forward for compression train and trunk pipelines respectively, in order to confirm whether the designs of base case are optimal. Otherwise, optimal options would be applied for an optimal case. Here the 'optimal case' is not strictly derived from optimization study, just by comparing several options. Finally, the overall costs of the whole pipeline network in the base case and the optimal case are summarized and compared. Figure 3 shows the work flow of the techno-economic evaluation in this study.

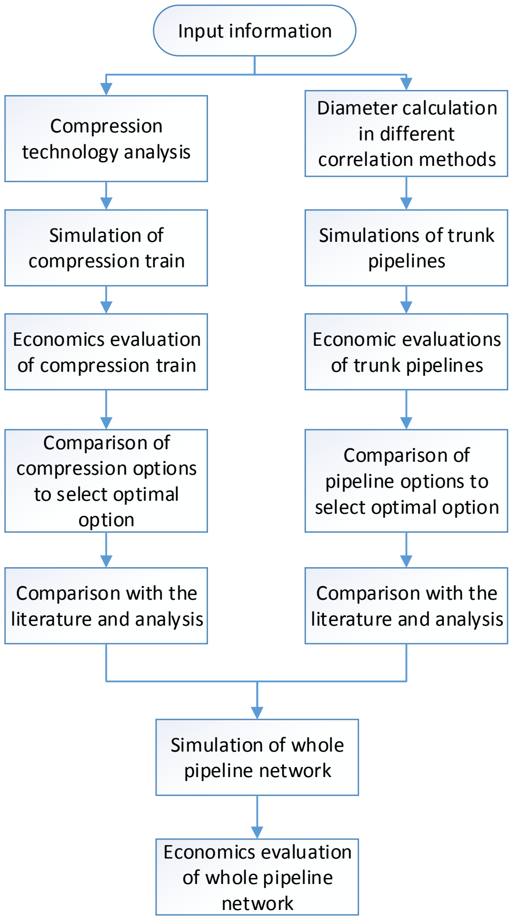

Figure 3. The work flow of the techno-economic evaluation

4. Techno-economic evaluation of CO_2 compression

Various types of compression configurations for CO_2 pipeline transport were found in literature. In the study of Zhang et al. [9], 5 stage centrifugal compression was applied for pressurization power consumption analysis. The exit temperature of the intercoolers is 20 °C in the subcooled liquid phase case. McCollum and Ogden [29] evaluated the energy cost, CAPEX and O&M cost of the compression achieved with 5 stage centrifugal compression followed by pumping. Witkowski et al. [56] performed a thermodynamic evaluation of various CO_2 compression configurations and only the power requirements of those options were compared. In this section, the case studies about the compression train at the White Rose plant was conducted to get optimal configuration. The results were compared with other published studies in literature.

4.1. Compression configuration options

After the pre-treatment process, the CO_2 captured in the White Rose plant will be pressurized from 1bar to 136 bar for dense phase transport by a compression train. The process model in this section only includes the inlet/outlet CO_2-rich streams and compression train. Four different compression configurations (see Table 6) were selected for techno-economic evaluation and compared with the base case (see Table 3). For options C3 and C4, the CO_2 mixture is initially pressurized to supercritical pressure (80 bar considering the impurities content in this study) and then further pressurized to the final exit pressure 136 bar by pumping. The difference between option C3 and C4 is the exit temperature of the intercoolers.

Table 6. Compression technology options and their process definition

Option	Unit	Base case	C1	C2	C3	C4
Description		Centrifugal 5 stages with 4 intercoolers	Centrifugal 16 stages 4 intercoolers	8 stages centrifugal geared with 7 intercoolers	6 stages integrally geared with 5 intercoolers to 20 °C +pumping	6 stages integrally geared with 5 intercoolers to 38 °C +pumping
Capacity	t/h	334.596	334.596	334.596	334.596	334.596
Suction pressure	bar	1	1	1	1	1
Suction temp.	°C	20	20	20	20	20
Pumping suction pressure	bar	-	-	-	80.0	80.0
Pumping suction temp.	°C	-	-	-	20	20
Exit pressure	bar	136.0	136.0	136.0	136.0	136.0
Stage	-	5	16	8	6	6
Isentropic efficiency	%	75	75	75	75	75
Intercooler exit temperature	°C	20	38	38	20	38
Last stage exit temp.	°C	20	20	20	20	20

4.2. Results and analysis

The comparison of energy and utilities requirement for the five compression configurations can be seen in Table 7. Option C3 has the lowest annual energy and utilities cost. The intercooling performance is one of the key factors related with the energy consumption of the compressor. Option C2 has less compressor stages but more intercoolers than option C1. Option C2 has 12.78% annual energy and utilities cost saving compared to option C1 and the most is energy saving. Compared to option C4, the lower intercooler exit temperature in option C3 (20 °C vs. 38 °C) between each stage results in 3.10% energy saving and 4.59% saving in annual energy and utilities cost. Option C3 has 5.44% energy saving, but only 2.07% annual energy cost saving compared with option C2. The reason is that in option C3 a suction temperature of 20 °C is specified for the pumping to cool down the CO_2 mixture with suction pressure at 80 bar, to avoid any gas formation for the pumping, which cause higher refrigerant cost.

Table 7. Energy and utilities requirements of compression technologies

Option	Energy requirements (kWh)	Cooling duty (m^3/h)	Refrigerant (t/h)	Energy and utilities cost (M€/a)
Base case	34546	-	1257	23.13
C1	39921	2540	656	26.29
C2	34832	2977	423	22.93
C3	31921	-	1197	21.42
C4	32972	2304	592	22.45

Figure 4 shows the comparison of levelized cost in breakdown of these five compression configurations. The range of total levelized costs is from 11.81 €/t-CO_2 to 14.99 €/t-CO_2. Energy and utilities cost is the biggest part with a proportion range of 65.6 % to 71.3%. Option C3 has the lowest total levelized cost of 11.81 €/t-CO_2 although levelized capital cost of option C3 is 2.14 €/t-CO_2, 0.25 €/t-CO_2 higher than the base case. The reason is that lower pressure ratio of each stage compression benefits a big saving of energy and utilities consumption. Compared with the base case, option C3 has an annual saving of 1.13 M€.

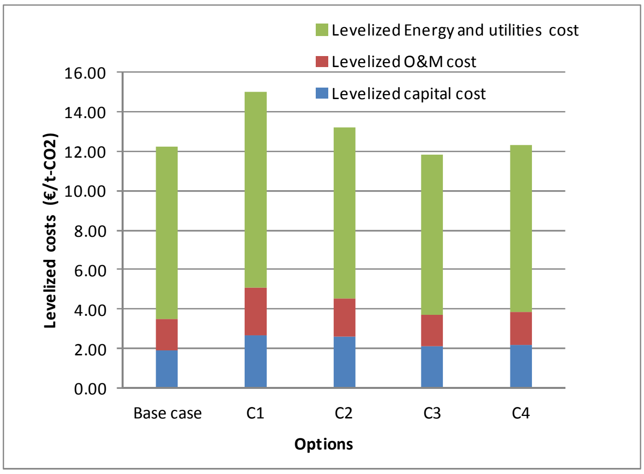

Figure 4. Comparison of levelized costs of different compression options

4.3. Comparison with other studies in the literature

There is a little published literature about the cost estimate of CO_2 compression. IEA-GHG [27] proposed an equation for the calculation of capital cost of compression based on the power required. Ogden et al. [31] developed a correlation summarizing the data from Carbon Capture Project (CCP). The annual O&M cost was calculated by applying a factor of 0.04 to the total capital cost. Wong [57] reported that the typical levelized cost of CO_2 compression varies from 5.5 € to 7.4 € per tonne of CO_2 with an estimated capital cost of 4.12 M€ per 3000HP in average. The method for O&M calculation was not mentioned in the paper. McCollum and Ogden [29] studied the cost of the compression train with 5-stage compression followed by pumping and the O&M factor is also 0.04.

For the energy and utilities cost, it is generally accepted that it can be accurately calculated based on the consumption data of process simulation results. So it was not included in the comparison. Figure 5 shows the comparison of levelized capital cost and O&M cost of different cost models used in IEA-GHG [27], McCollum and Ogden [29]and this study. The method used by McCollum and Ogden [29] failed to distinguish the costs of different options as a flow-based equation was applied for the capital cost calculation. The comparison shows the O&M cost in this study is much higher than in other two methods.

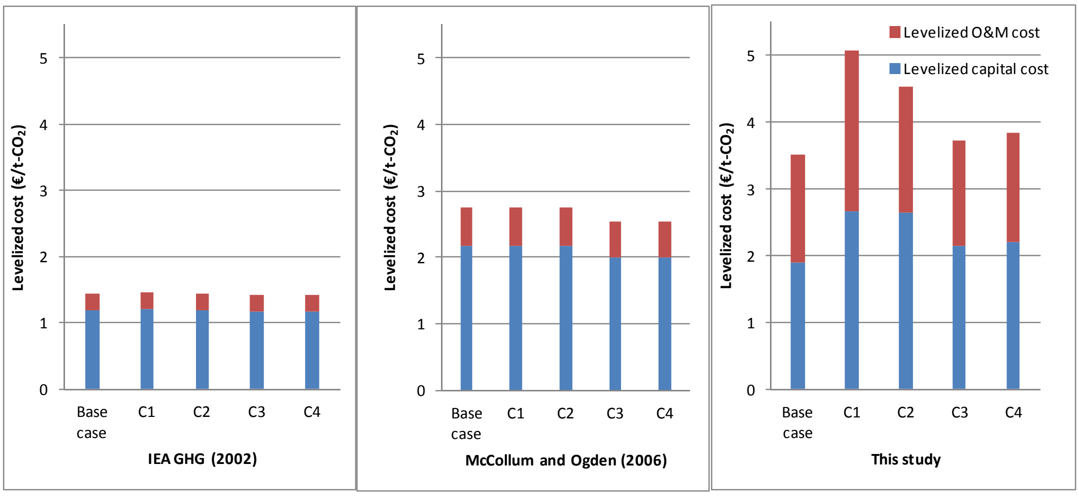

Figure 5. The comparison of levelized cost of different cost model

5. Techno-economic evaluation of trunk pipelines

In section 2, the diameters of the onshore and offshore trunk pipelines were selected with a velocity-based equation for the base case. In this section, different published pipeline diameter models were used for diameter calculation and different results were obtained. Steady-state simulations were conducted to do a rating calculation for different diameters in order to compare process performance and economic evaluation. The model used in this section only includes the trunk pipelines and the booster pump station. As the same calculation base for each simulation model (see Table 8), the input conditions of the trunk pipelines are the mixture of the CO_2-rich streams coming from two power plants (see Table 3). The results of different models were compared and the optimal diameter was chosen for the optimal design.

Table 8. Input and boundary conditions

Condition	unit	Value
Composition of CO_2 mixture stream	mole%	96%CO_2, 2%N_2, 1.41%H_2, 0.59%Ar
Flow rate	t/h	1126.263
Entry pressure	bar	136
Entry temperature	^oC	20
Minimum arrival pressure at offshore platform	bar	126

5.1. Calculation of pipeline diameter

The diameter is a key factor for both technical and economical assessments in designing a pipeline system. For a given CO_2 pipeline transport task, different published models can be used to calculate the diameter of the pipelines. Table 9 shows an overview about the equations of several published models. The velocity based equation is often used to do an initial estimation by setting input velocity in an experienced economical range. The (extensive) hydraulic equation is only capable for the liquid transport. McCoy and Rubin model can be used for both gaseous and liquid phase transport because it integrates the equation of state of real gas with the energy conservation and hydraulic equations.

Table 9. Overview of the different diameter calculation methods in literature

Name	Formula	Abbreviations	Source
Velocity based equation	$D = \sqrt{\frac{4m}{\upsilon \pi \rho}}$	D =diameter (m), m =mass flow (kg/s), υ =velocity (m/s), ρ=density (kg/m^3)	Wildenborg et al., 2004; Element Energy, 2010; Chandel et al., 2010
Hydraulic equation	$D = \left(\frac{32 f m^2 L}{\pi^2 \rho \Delta P}\right)^{1/5}$	f =Fanning friction factor, L=length (m), ΔP=overall pressure drop (Pa)	Heddle et al. [24]); Van den Broek et al. [23])
Extensive hydraulic equation	$D = \left(\frac{4^{10/3} n^2 m^2 L}{\pi^2 \rho^2 (\Delta h + (\Delta P/\rho\, g))}\right)^{3/16}$	n =Manning friction factor, Δh =height diffirence (m), g=gravity constant (9.81m/s^2)	Piessens et al., 2008
McCoy and Rubin model	$D = \left(\frac{-64 Z_{ave}^2 R^2 T_{ave}^2 m^2 f L}{\pi^2 [M Z_{ave} T_{ave} R (P_2^2 - P_1^2) + 2 g P_{ave}^2 M^2 \Delta h]}\right)^{1/5}$	Z_{ave} =Average fluid compressibility, R =Gas constant (8.31Pa M^3/mol K), T_{ave}=average fluid temperature (K), M =molecular weight of fluid (kg/kmol), P_1 =Pressure at inlet (Pa), P_2 =Pressure at outlet (Pa), P_{ave} = Average pressure in the pipeline= $\frac{2}{3}\left(P_2 + P_1 - \frac{P_2 \times P_1}{P_2 + P_1}\right)$	McCoy and Rubin, 2008; Gao et al., 2011

For the diameter calculation, the parameters of the CO_2 mixture stream were obtained from the process simulation results and are substituted into each equation. Table 10 presents the results of calculated diameter of trunk pipelines. For the velocity based method, 1.0m/s, 1.5 m/s and 2.0m/s were selected for the diameter calculation. The results show the velocity range of other three methods is from 1.3 to1.8 m/s, which is close to the most effective velocity range of from 1.5 to 2.0 m/s [25]. As only standard size pipeline diameters (ANSI standard) are specified in APEA, the calculated diameters were rounded off to the nearest whole number. With a diameter of 20 inches, the pressure somewhere of the onshore trunk pipeline is below 101 bar, which does not meet the operational constraint. The diameters of 22, 24 and 28 inches were then selected for the techno-economic evaluations in next section.

Table 10. The calculation results of different diameter models

Item Unit	Calculated diameter (m)	Velocity (m/s)	Selected diameter in APEA DN (inch)
	0.699	1.0	28
Velocity based equation	0.5713	1.5	24
	0.4948	2	20
Hydraulic equation	0.5262	1.77	22
Extensive hydraulic equation	0.6173	1.29	24
McCoy and Rubin model	0.5672	1.52	22

5.2. Results and analysis

The selected diameters were used as the inputs in steady-state models in order to simulate the hydraulic performance of the pipeline. The results of each simulation were exported into APEA to do the economic evaluations. Table 11 shows hydraulic results and power requirement of each simulation. Higher velocity results in a greater pressure drop of the CO_2-rich stream in the onshore and offshore trunk pipelines. Higher boosting pressure of the pump station is then needed to compensate the pressure loss to maintain a constant arrival pressure at the offshore storage platform.

Table 11. Hydraulic performance and energy requirement of trunk pipelines in different diameters

Pipeline diameter (inch)	Actual initial velocity (m/s)	Pressure drop of onshore pipeline (bar)	Pressure drop of offshore pipeline (bar)	Boosting pressure of pump station (bar)	Energy required of pump station (kWh)
28	1.08	5.9	10.0	5.9	301.5
24	1.49	13.5	20.6	24.1	1243
22	1.81	22.1	32.2	44.3	2305

Fig. 6 illustrates the comparison of the levelized cost in breakdown of three options with different diameters. The comparison shows that the saving of capital cost is much bigger than the penalty of energy cost when the diameter of the pipelines decreases from 28 inches to 24 inches and then to 22 inches. The option with 22-inch diameter has the lowest total levelized cost of 7.59 €/t-CO_2. Compared with the option of 24-inch diameter in the base case, the option with 22-inch diameter has an annual saving of 7.34 M€.

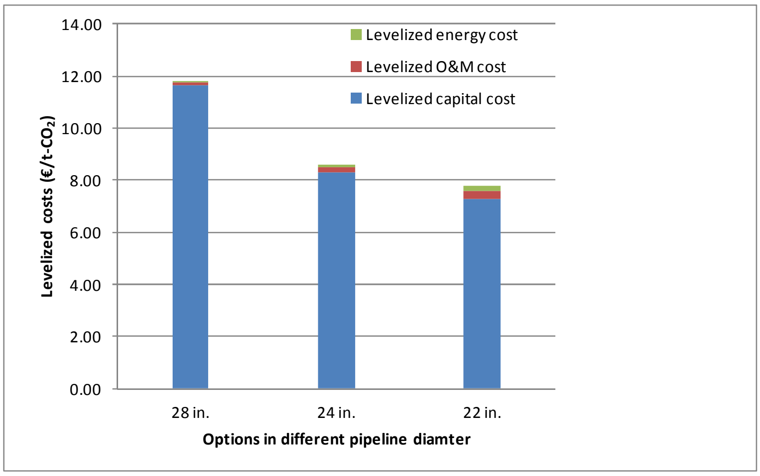

Figure 6. Levelized cost comparison for different diameters of the pipelines

5.3. Comparison with other studies in the literature

For the published models [23-29, 58] for cost evaluation of the CO_2 pipelines, some of them do not include a cost assessment of the booster pump. None of them can make an economic evaluation of the pipelines integrated with the energy cost of booster pump station. For comparison, the capital costs of trunk pipelines were calculated respectively by different methods developed by IEA-GHG [27], McCollum and Ogden [29], Piessens et al. [58] and Van den Broek et al. [23]. Figure 7 shows a large range of the capital cost per kilometre of pipeline from estimating by different cost models. The total capital cost calculated in this study and Piessens et al. [58] is much higher than those calculated with the other models. The main reason is that the method used by Piessens et al. [58] and in this study is based on the weight of material while the other methods are mainly based on the historical cost data of natural gas pipelines. Those correlation models, except for the weight-based models, do not consider the adaptation for the higher operation pressure of CO_2 pipeline transport. Normally, higher design pressure requires higher wall thickness of the pipelines, which results in a significant increase of the material cost.

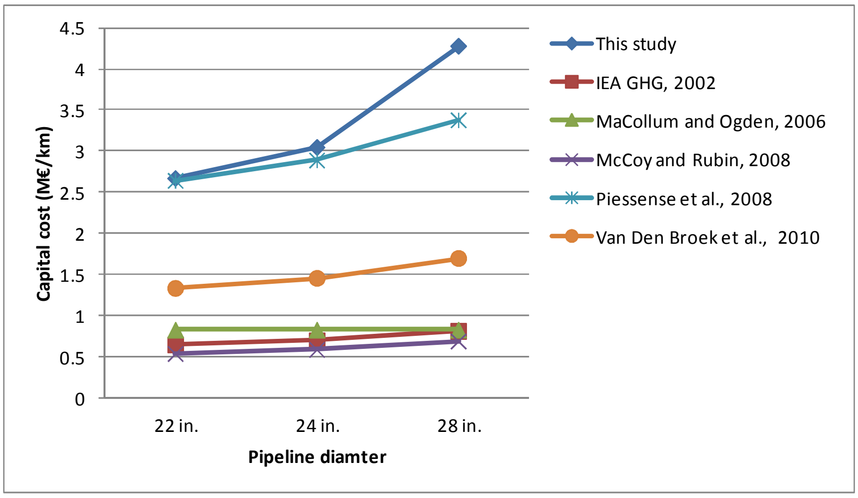

Figure 7. Comparison of capital cost of different cost models

6. Overall cost of CO_2 transportation pipeline network

6.1. Comparison of the base case and the optimal case

In sections 4 and 5, techno-economic evaluations were conducted for the compressors and trunk pipelines respectively. The options, which have the lowest annual costs, were used to optimize the design of the pipeline network in this study. For the potential configuration of compression train at the White Rose plant, amongst five options compared in Section 4, C3 option (see in Table 6) is the optimal option. The compression train for the Don Valley plant could apply the similar configuration but it includes two parts, 5-stage compression and 1-stage compression followed by pumping as the CO_2 mixture will be transported in the gaseous phase at a pressure of 35 bar first and then boosted to dense phase at a pressure of 136 bar before entering the trunk pipelines. For the trunk onshore and offshore pipelines, 22-inch diameter is the optimal option (refer to Figure 6). Then the cost of the optimal case of whole pipeline system can be calculated. As mentioned in Section 3, the ‘optimal case’ is not strictly derived from optimization study, just by comparing several options in this study.

The overall cost of the base case and the optimal case were compared in Figure 8. The total capital cost was split into the costs of trunk pipeline and collecting system for a better comparison. The collecting system includes the collecting pipelines and compression trains.

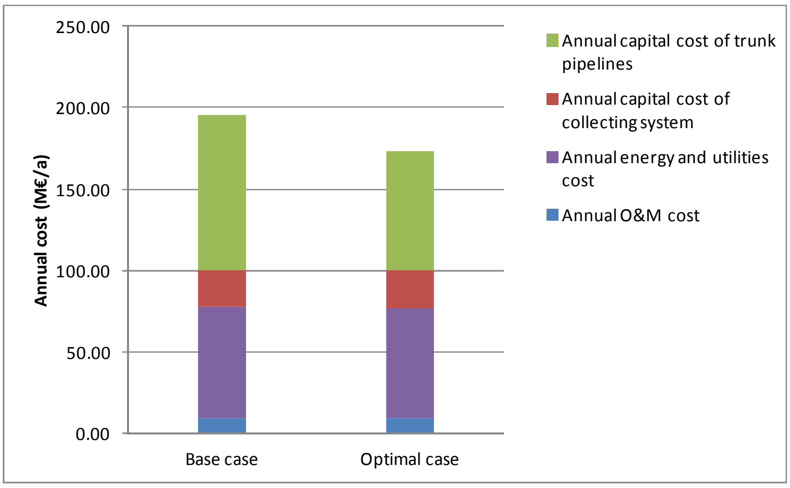

Figure 8. Comparison of annual costs of base case and optimal case

The comparison shows the annual O&M cost is almost same for two cases as they are similar processes. The annual energy and utilities cost is also very close to each other. Table 7 in Section 4 shows a significant compression energy saving for the optimal case compared to the base case. However the smaller diameter of trunk pipelines in optimal case increased the frictional pressure drop along the pipelines. This requires higher boosting pressure and therefore, higher energy consumption of the booster pump. The annual capital cost of trunk pipelines of the optimal case is obviously lower than the base case. Smaller diameter of pipelines has the advantage of incurring lower material cost and the construction cost may also be lower. Compared to the base case, optimal case has an annual total saving of 22.7 M€. But it should be noticed that, having a larger diameter trunk pipelines, the base case provides the opportunities to transport extra CO_2 from additional electricity generation capture plants or industrial capture plants in this region in the future.

6.2. Comparison with other studies in the literature

Public data are scarce for a cost comparison about the whole pipeline network for CO_2 transport. Most published studies present the costs evaluation for the pipelines without including a cost assessment of the compression train. The few studies that carried out an evaluation of the compression train failed to link it to the whole pipeline network system. In the study of Roussanaly et al. [59], the economic evaluations were conducted to compare different options for the COCATE project. The cost evaluation of the onshore pipelines option presented a typical pipeline network comprising a collecting pipeline system (including compression) around 40 km long and an onshore trunk pipelines around 620 km long.

The levelized costs per tonne of CO_2 were summed up in each of the studies as shown in Figure 9. The levelized energy and utilities cost is close for these two studies. The levelized capital cost of trunk pipelines for the COCATE project is about 5.5 €/t-CO_2 , much lower than 8.1 €/t-CO_2 of the optimal case in this study, despite the fact that length of the COCATE pipeline is 620 km while the length of pipeline used in this study is 162 km. The evaluations of COCATE project used a specific pipeline cost model based on pipeline data of several published cost models. The reason for low capital costs predicted by most of the published models was analyzed in section 5. The levelized capital cost of collecting system in COCATE project is only 0.2 €/t-CO_2. The details of the evaluation method used for the collecting pipeline system in the COCATE project were not reported in the paper.

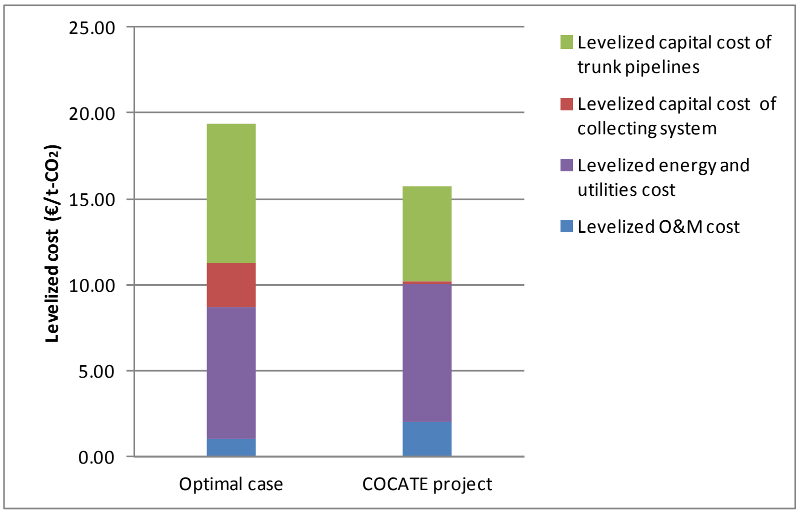

Figure 9. Comparison of levelized cost of the optimal case and COCATE project

7. Conclusions

This paper is aimed to carry out simulation-based techno-economic to explore the optimal design of the CO_2 transportation pipeline network, in order to reduce the cost of CCS deployment. A detailed steady state model was developed for transport system comprising CO_2 mixture streams from two emitters, the compression train, the onshore and offshore trunk pipelines and the booster pump station. The simulation results were exported into APEA for conducting the techno-economic evaluations. The optimal options with the lowest annual cost for compression train and trunk pipelines were selected after comparative studies of the different options. The overall costs of base case and optimal case were also compared. The optimal case has an annual total saving of 22.7 M€ compared with the base case. For the optimal case, levelized energy and utilities cost is 7.62 €/t-CO_2, levelized capital cost of trunk pipeline is about 8.11 €/t-CO_2 and levelized capital cost of collecting system is 2.62 €/t-CO_2. The cost evaluation results of the compression train, trunk pipeline and whole pipeline network were compared with the cost evaluation results in open literature respectively to gain more insights. More conclusions are seen as below:

- For CO_2 compression, lower intercooler exit temperature (20 °C vs. 38 °C in this study) and lower pressure ratio per stage results in lower energy and utilities consumption of compression train.

- The correlation based cost models for CO_2 compression train cannot give good cost predictions for some different configuration options. The O&M factor of 0.04 used in those models is very small compared with the result of this study.

- The pipeline diameter models in the literature are generally reliable. The initial velocity of CO_2 –rich stream is around 1.7m/s in the optimal case in this study.

- A large range of capital cost was obtained after applying different published cost models for the trunk pipelines. Compared with the results in this study, most of the pipeline cost models in the literature predicted a much lower capital cost and the weight-based model in the study of Piessens et al. [58] has the closest prediction with this study.

- Simulation-based techno-economic evaluation method offers a powerful tool for decision making support for selecting detailed technical options.

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