CASE STUDY ON CO_2 TRANSPORT PIPELINE NETWORK DESIGN FOR HUMBER REGION IN THE UK

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ABSTRACT

Reliable, safe and economic CO_2 transport from CO_2 capture points to long term storage/enhanced oil recovery (EOR) sites is critical for commercial deployment of carbon capture and storage (CCS) technology. Pipeline transportation of CO_2 is considered most feasible. However, in CCS applications there is concern about associated impurities and huge volumes of high pressure CO_2 transported over distances likely to be densely populated areas. On this basis, there is limited experience for design and economic assessment of CO_2 pipeline.

The Humber region in the UK is a likely site for building CO_2 pipelines in the future due to large CO_2 emissions in the region and its close access to depleted gas fields and saline aquifers beneath the North Sea. In this paper, various issues to be considered in CO_2 pipeline design for CCS applications are discussed. Also, different techno-economic correlations for CO_2 pipelines are assessed using the Humber region as case study. Levelized cost of CO_2 pipelines calculated for the region range from 0.14 to 0.75 GBP per tonne of CO_2. This is a preliminary study and is useful for obtaining quick techno-economic assessment of CO_2 pipelines.

Keywords: *Case Study, CO_2 transport, Pipeline, Network Design, Techno-economic analysis*

NOMENCLATURE

A	Cross sectional area (m^2)
C	Conversion factor
C_{annual}	Annual cost (GBP)
D	Internal diameter (m)
D_{opt}	Optimal inner diameter (m)
E	Seam joint factor
f	Friction factor
F	Safety factor
j	Discount rate (%)
n	Operational lifetime (years)
Q_m	Mass flowrate (kg/s)
Q_s	Gas volumetric flowrate at standard condition (Sm^3/d)
L	Length of pipe (m)
P	Pressure (Pa)
P_{gas}	Pressure (Gaseous phase condition) (Pa)
P_{liquid}	Pressure (Liquid phase condition) (Pa)
$P_{in,}P_{out}$	Inlet and outlet pressure of a pipe section (Pa)
P_{max}	Max. design pressure (MPa)
S	Pipe yield strength (MPa)
t	Pipe thickness (m)
T	Temperature (K)
Z	Average gas compressibility factor

Greek Alphabets

Δ	Change in
ε	Roughness (m)
μ	Dynamic viscosity (kg/m.s)
π	Pie
ρ	Mass density (kg/m^3)
ρ_s	Gas density at standard condition (kg/m^3)
v	Flow velocity (m/s)

1 INTRODUCTION

1.1 Background

Reducing CO_2 emissions from industries and the power sector is necessary for achieving stable atmospheric concentrations of CO_2 [1]. CCS is recognised to have the potential of economically and reliably meeting emission reduction expectations from these sources [1,2]. CCS involves separation of CO_2 from process effluents/flue gases to obtain concentrated CO_2 stream which is thereafter compressed to high pressures and transported to geological storage/EOR sites where they are injected and prevented from entering the atmosphere. Sustainable, safe and economic CO_2 transport is vital for realizing CCS technology. IPCC [1] and Svensson *et al.* [3] showed that pipeline transportation is the most feasible transportation option. This is due to large volumes of CO_2 expected to be transported over long distances. Pipeline transport of CO_2 is similar to the pipeline transport of hydrocarbons. However, design and operation of CO_2 pipeline is more complicated due to the highly non-linear thermodynamic properties of CO_2 and transportation of CO_2 at pressures above the critical pressure (dense phase) [4].

High pressure CO_2 pipelines have been in operation in North America since 1972 and globally there are about 5800 km for high pressure CO_2 pipelines transporting about 50 Mt/year of CO_2 for EOR applications [1]. The CO_2 transported are obtained from naturally occurring sources (Cortez, Sheep Mt, Bravo, Central Basin pipelines) and gasification plants (Canyon Reef, Weyburn, Val Verde, Bairoil pipelines) at fairly pure condition [1]. The CO_2 pipelines in these applications are routed through sparsely populated areas. The experience from these applications can be transferred to CO_2 pipeline transport for CCS.

For CCS applications, the CO_2 will mostly be obtained from fossil fuel power plants and carbon-intensive industries. Unlike existing EOR CO_2 pipelines where the CO_2 supply to the pipeline is fairly steady, CO_2 supply to the pipeline network are subject to load levels in the power plant and will consequently change continuously. The CO_2 stream also contains impurities and will likely be routed through densely populated areas. The amounts of impurities are subject to the technology deployed for capturing CO_2; post-combustion, pre-combustion and oxy-fuel CO_2 capture [1]. Details of the different capture technologies are available in Wang *et al.* [5]. **Table 1** shows typical impurity levels in a CO_2 streams captured using different technologies.

Table 1: Concentration of impurities in dried CO_2, expressed in % per volume [1]

	SO_2	NO	H_2S	H_2	CO	CH_4	$N_2/Ar/O_2$	Total
COAL-FIRED PLANTS								
Post-combustion capture	<0.01	<0.01	0	0	0	0	0.01	0.01
Pre-combustion capture	0	0	0.01-0.6	0.8-2.0	0.03-0.4	0.01	0.03-0.6	2.1-2.7
Oxy-fuel	0.5	0.01	0	0	0	0	3.7	4.2
GAS-FIRED PLANTS								
Post-combustion capture	<0.01	<0.01	0	0	0	0	0.01	0.01
Pre-combustion capture	0	0	<0.01	1.0	0.04	2.0	1.3	4.4
Oxy-fuel	<0.01	<0.01	0	0	0	0	4.1	4.1

1.2 Stationary Emission Sources of CO_2 in the UK

Industries and the power sector are the largest stationary emitters of CO_2 (>0.1 $MtCO_2$ per year) and CCS deployment are targeted at these sectors [1]. In the UK, they contributed $\sim 10\%$ and $\sim 32\%$ respectively of total CO_2 emissions in 2009 [6]. A geographical spread of top CO_2 emitters in the UK (power plants and industrial processing plants) are grouped into five clusters as shown in **Figure 1.** Due to proximity of the CO_2 sources within each cluster and large amount of emissions, the clusters are identified as first candidates for roll out of CCS technology.

The Humber region contributes the largest CO_2 emission (about 23% of total CO_2 emission in the UK [7]) and it is close to storage/EOR sites located beneath the southern North Sea (off the east coast of England) [8]. CCS development in the region could reduce UK's total current CO_2 emission by around 10% [9]. As a result, CO_2 pipeline development for this region is considered in this paper.

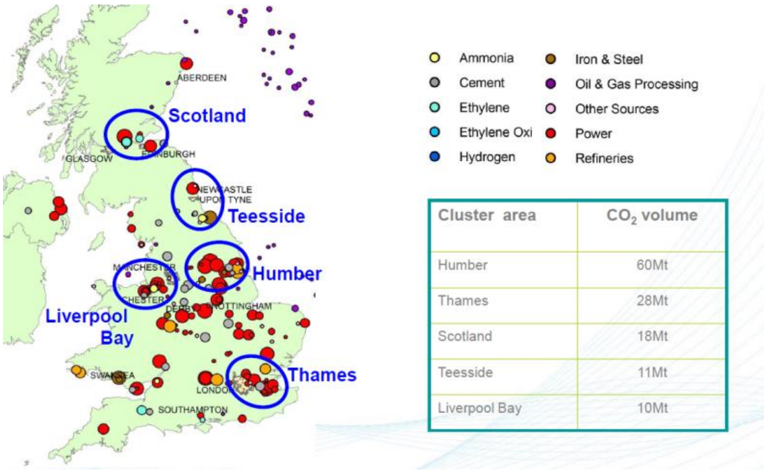

Figure 1: UK CO_2 Emissions clusters [10]

1.3 Motivations

CO_2 pipelines design and operation for CCS differs from that of natural gas transport pipelines. This is due to the complexities introduced by the presence of impurities, thermodynamic behaviour of CO_2, higher operating pressure (above CO_2 critical pressure), health and safety among others. As such, detailed techno-economic studies of CO_2 pipelines for CCS are required.

Also, it has become urgent that techno-economic assessment of CO_2 pipelines be conducted using definite case studies. As noted in **Section 1.2**, the Humber region in UK is strategic for CCS deployment. As such, the region constitutes a good basis for techno-economic assessment of CO_2 pipelines for CCS applications.

1.4 Aim and Novelties

The paper presents an assessment of techno-economic options for CO_2 pipeline network for the envisioned Humber region CCS hub in the UK. Related studies by Seevam *et al.* [11] focuses on assessing the suitability of re-using existing pipeline structures for CO_2 transport. This option is however subject to pressure ratings of the existing pipelines, the age of the pipelines and extent of corrosion. Contrary to that, this paper considers techno-economic assessment of new pipelines.

This paper also differs from other techno-economic studies of CO_2 pipelines [12-16] by providing an assessment of different techno-economic models for CO_2 pipeline network based on the Humber region case study.

2 GENERAL CONSIDERATIONS FOR CO_2 PIPELINE DESIGN

In developing suitable pipeline design, considerations are given to pipeline integrity, flow assurance, operation and health and safety. This will involve determining physical properties of the process stream, optimal size and pressure rating of the pipeline, examination of the topography of pipeline channel, geotechnical considerations and the local environment [1]. Some of these considerations are discussed further below.

2.1 Physical States of CO_2

Depending on temperature and pressure, CO_2 can exist in different states: gas, liquid, solid and supercritical/dense phase (**Figure 2**). The figure further shows:

- The 'triple point' at (5.2 bar, -56°C), where CO_2 can exist in all three phases,
- The 'critical point' at (73.76 bar, 30.97°C), above which CO_2 exists in supercritical phase and
- The supercritical region found beyond critical pressure and temperature. Dense phase condition is obtained above critical pressure but below critical temperature. At supercritical/dense phase condition, CO_2 have density closer to the liquid phase and viscosity similar to the gaseous phase.

Supercritical/dense phase is the best phase for pipeline CO_2 transport due to the high density and low viscosity [3,17-21]. Pipeline transportation in the liquid phase is also possible [22]. CO_2 liquefies at temperatures below the critical temperature and pressures above 516.7kPa but less than the critical pressure. At this condition, they are more suitably transported through vessels like LNG and LPG [1]. Gaseous phase pipeline is not economical due to low gas density and consequently high pressure drops [22]. Larger pipes will also be required for transporting equivalent amounts of CO_2 when they are in gaseous phase. However, where CO_2 is to be reused in gaseous phase at relatively short distances, transport in gaseous phase may be reasonable.

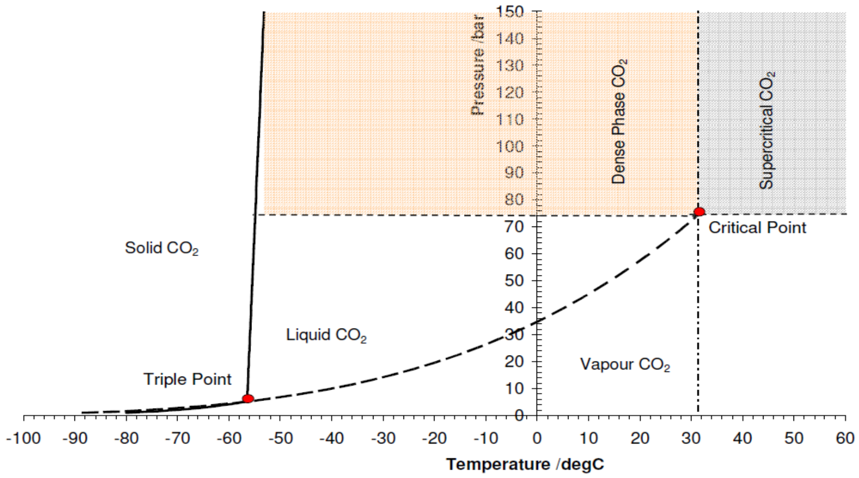

Figure 2: Phase diagram for pure CO_2 [23]

When transported in the supercritical/dense phase condition, the pressure should be maintained above the critical pressure to prevent formation of liquid, gas or two-phase flow condition. Two-phase flow can cause operational problems such as slugging and cavitations in pumps/compressors. In practice, CO_2 will have to be compressed to a high pressure at the starting point and re-pressurized in booster stations along the pipeline to make up for head losses and maintain the required level of pressure above the critical point. Operating at such high pressure will require thicker pipelines that can contain the pressure. At dense phase condition, conventional seals used in valves, pumps and compressors tend to absorb CO_2 and could easily become embrittled when depressurised. These are key concerns for adopting existing oil and gas pipelines for CO_2 transport.

CO_2 have large values of Joule-Thompson coefficient. In the event of rapid depressurization due to leakage for instance, its rapid expansion from compressed state is consequently accompanied by significant cooling effect. This could be to the extent that solid CO_2 (“dry ice”) is formed. This has health and safety implications.

2.2 Density and Viscosity

Density and viscosity are important variables in the design of CO_2 pipelines. They are strongly dependent on temperature and pressure. Pressure drop along the line will reduce the density and increase velocity of CO_2 stream which will in turn increase the pressure drop (**Figure 3**). At two-phase flow condition which is possible below critical condition, sudden change in density results leading to increased pressure drops and greater energy expenditure [14]. This can be seen in **Figure 3** where density decreases with pipeline length reflecting pressure drop mostly due to friction.

High pressure drops resulting from increasing flow velocity leads ultimately to “choking” condition [22]. This can be prevented by pressure boosting at some pipeline length 10% less than the choking distance. Booster stations will however add to cost of transport. Two-phase flow can also result in hilly terrain where low pressure points can arise within the pipeline.

Changing temperature also affects density of CO_2 [18]. However, temperature changes are not a major problem because the pipelines are buried at depths with fairly stable temperature of the surroundings. CO_2 viscosity behaves similarly with changing temperature and pressure.

CO_2 is denser than air. In the event of leakage, they may settle at low lying regions. This is a major health and safety concern and requires developing suitable response strategy in the event of leakages.

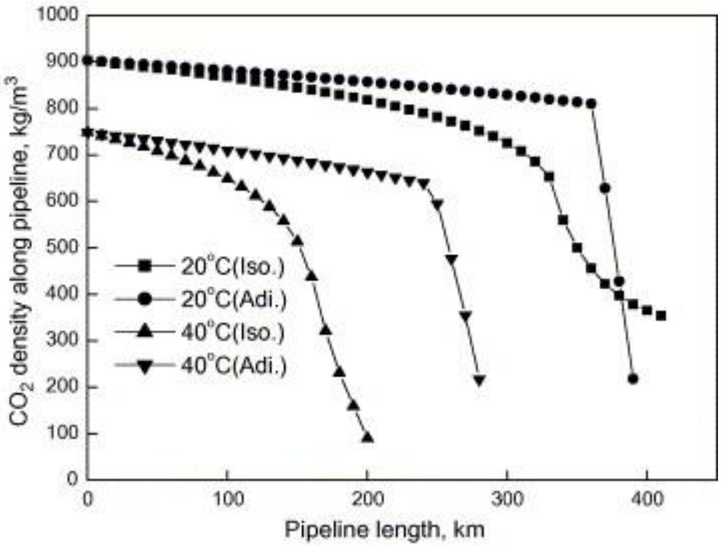

Figure 3: CO_2 density changes along pipeline at different inlet temperatures [22]

2.3 Impurities in CO_2 Stream

Impurities in the CO_2 stream (**TABLE 1**) alter the physical properties of the transported CO_2 to an extent that will depend on the individual concentration of the impurities [14].

2.3.1 Effect of impurities on critical properties of CO_2

SO_2 and H_2S impurities increase the critical temperature of the mixture, while CH_4, Ar, N_2, or O_2 lowers the critical temperature of the mixture [21] **(Figure 4).** On the other hand, SO_2 and N_2 lower the critical pressure of CO_2 mixtures while the presence of H_2S, CH_4, Ar, and O_2 in the mixture has negative influence on critical pressure **(Figure 4).** In addition, NO_2 impurity is reported to increase both critical temperature and pressure when present in binary or ternary mixtures with CO_2 [18]. Higher critical temperature and lower critical pressure will be beneficial for pipeline transport since less energy will be used during compression [21].

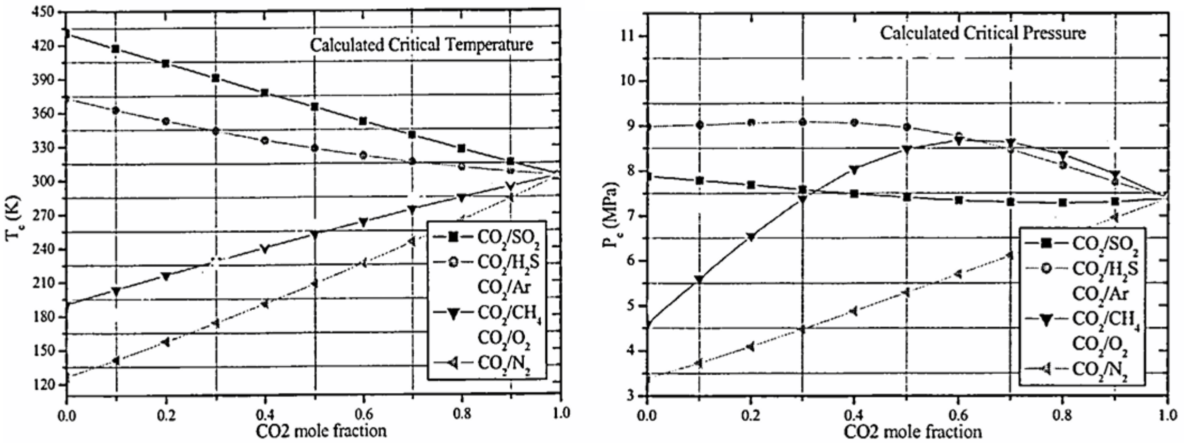

Figure 4: Critical temperature and pressure of binary mixtures with CO_2 [21]

2.3.2 Effect of impurities on phase envelope of CO_2

The phase behaviour of CO_2 is affected by impurities as shown in **Figure 5**. Impurities increase the two-phase region; H_2 and NO_2 increase the area of two-phase envelope while N_2 and H_2S have much smaller influence [18]. This change increases the possibility of slug flow. CO_2 mixtures involving NO_2 are also shown to have phase envelopes below vapour-liquid line of pure CO_2. To maintain dense phase condition as expected, concentration of impurities should be as low as possible otherwise further energy expenditure will be incurred [1]. Other effects include boiling point rise by SO_2 or H_2S impurities and boiling point depression by CH_4, Ar, N_2, or O_2 impurities [21]. H_2S have the least effect on boiling point of the CO_2 stream [21].

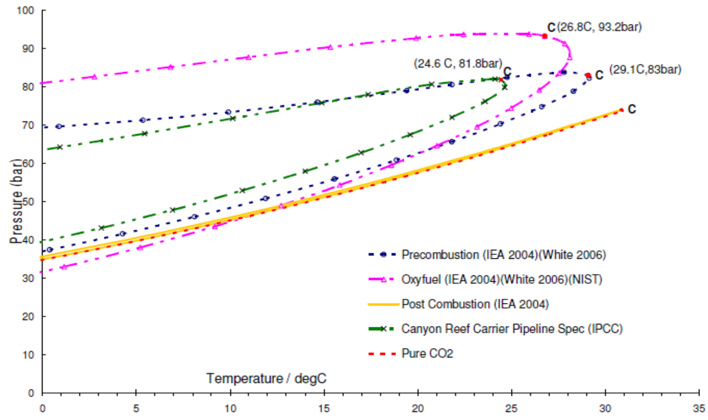

Figure 5: Phase diagram for the three different capture streams and the Canyon Reef Carrier pipeline specification [18]

2.3.3 Effect of impurities on density

Presence of impurities such as SO_2 with molecular weights higher than that of CO_2 results to increase in the CO_2 mixture density [21]. On the other hand, impurities with lower molecular weight (H_2S, CH_4, Ar, O_2 and N_2) lower the mixture density. Effect on the mixture density is most obvious with SO_2 and CH_4 impurities which respectively have the highest and lowest molecular weight of all the key impurities. Impact of impurities on density profile change resulting from phase changes has been investigated by Seevam *et al.* [18]. With decreasing pressure, a sudden change of density in pure CO_2 is observed as result of crossing over the vapour-liquid line. Due to different boiling and condensing temperatures of various CO_2 mixtures, effect of sudden change of density can happen at much higher pressure with two-phase conditions existing over wider range of pressures. In the mixture of 95%CO_2-5%H_2, density discontinuity effect happens at significantly higher pressures of 110-120 bar [18].

2.3.4 Effect of Impurities on Material of Construction

Free water impurities induce pipeline corrosion due to the formation of highly corrosive carbonic acid. Corrosion rate of 10 mm/year is quoted for a CO_2 pipeline with free water impurity [24]. It is further shown that corrosion rate increases with increasing temperature, higher flow velocity and in the presence of O_2 in the stream [25]. Corrosion risk is important in selection of pipeline material. When dehydrated to <100ppm, it is has been demonstrated that low alloy carbon steel can be used to transport high pressure CO_2 without significant corrosion issue [4]. Wet CO_2 streams will require pipeline made of a corrosion-resistant alloy material such as stainless steel [1]. However, this may not be economical.

Impurities such as hydrogen impurities could potentially diffuse into pipeline material [26]. This could embrittle the pipe material and reduce their ductile and tensile strengths. Fractures could propagate in the pipeline under this circumstance except appropriate mitigating measures are in place.

2.3.5 Miscellaneous Effects of Impurities

Impurities reduce the capacity of the pipeline compared to transport of pure CO_2 and increase compressor power consumption [18]. Capacity reduction can be as high as 25% for transport of CO_2 from oxy-fuel process through a pipe of 0.736 m (29 inch) in diameter. Other effects include higher pressure and temperature drop illustrated by CO_2 stream from oxy-fuel capture process which has the highest impurity level [18]. Hydrate formation at certain temperature and pressure in the presence of free water impurities is also likely. Hydrates cause serious operational and safety problems such as pipe blockage among others.

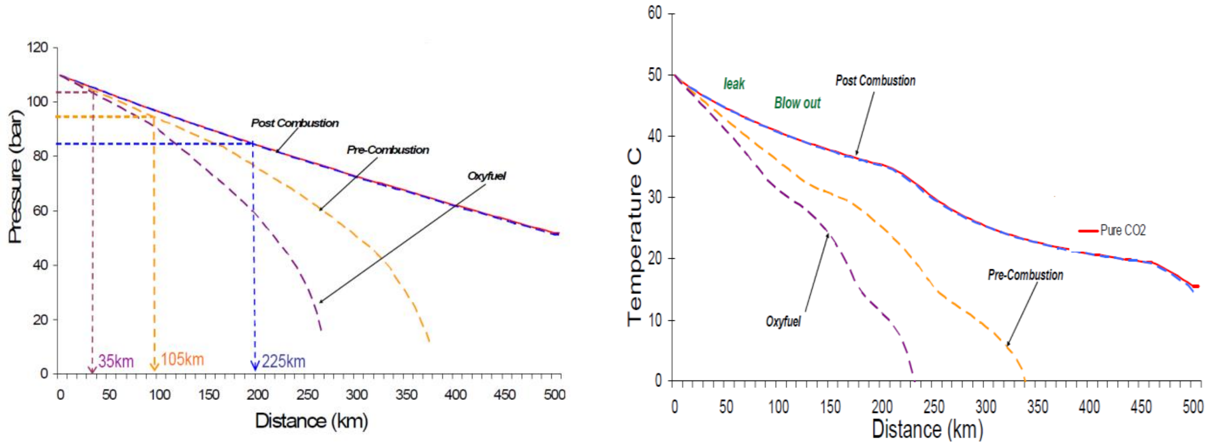

Figure 6: CO_2 Pipeline Temperature and Pressure Profile [18]

3 METHODOLOGY FOR TECHNO-ECONOMIC ANALYSIS

3.1 Technical aspect of CO_2 pipeline design assessment

The important parameters in pipeline flow capacity calculations are internal diameter, fluid density and viscosity, pipeline length, pressure drop over the pipeline, terrain profile, and temperature of both the surroundings and the fluid [17,19,20,22 27,28].

3.1.1 Diameter calculation

Diameter calculation can be obtained based on hydraulic laws or economic parameters [20].

3.1.1.1 Hydraulic methods

Existing hydraulic methods employ two different approaches – turbulent flow equations and equations with velocity as parameter [20].

Turbulent flow equations

$$D^5 = \frac{32fLQ_m{}^2}{\rho\pi^2\Delta P} \quad (1)$$

Equation 1 is derived from Bernoulli Equation for turbulent flow in a pipe neglecting kinetic energy and potential energy parameters [20]. The Colebrook equation or any convenient method is used to determine the friction factor, *f.* This method has been used successfully in CO_2 pipeline design by Zhang *et al.* [15].

Based on the hydraulic laws, alternative approach has also been proposed Vandeginste and Piessens [20] and Mohitpour *et al*. [29]. Their methods differ from the more general approach (**Equation 1**) because they account for elevation differences. In addition, Mohitpour *et al*. [29] method accounts for temperature. This approach offers more accurate prediction but require more information before they can be used.

<u>Equations with velocity as a parameter</u>

The method (**Equation 2**) is proposed by IEA GHG [30].

$$D = \sqrt{\frac{4Q_m}{\pi\rho v}} \tag{2}$$

3.1.1.2 Diameter estimation based on economic parameters

This is a cost optimal approach used by Zhang *et al*. [22].

$$D_{opt} = 0.363\left(\frac{Q_m}{\rho}\right)^{0.45}\rho^{0.13}\mu^{0.025} \tag{3}$$

This approach has limited accuracy. Smallest diameter (cost optimal diameter) will be accompanied by higher pressure drop and invariably higher pumping cost [20].

3.1.2 CO_2 Physical properties prediction

Density and viscosity of CO_2 are obtained using equations of state (EOS) at specific temperature and pressure conditions. Commonly used EOS includes Peng-Robinson, Redlich-Kwong, Soave-Redlich-Kwong, CSMA model, Span and Wagner equation among others [31]. Appropriate selection of EOS is important for accurate estimation of the physical properties and consequently good design of the pipeline network [21]. Properties of CO_2 may vary abnormally at dense phase conditions making accurate property estimation difficult when they are transported at this condition. Property estimation is further complicated by the presence of impurities.

Zhang *et al.* [22] used Peng-Robinson (PR) with Boston-Mathias modification to estimate CO_2 density and viscosity. PR show fairly accurate prediction at dense phase condition and can be applied to CO_2 mixtures [32]. Vandeginste and Piessens [22] used Span and Wagner equation. Span and Wagner cannot be applied to CO_2 mixtures. Chandel *et al.* [12] used the correlation proposed by McCollum and Ogden [19]. The correlation reportedly gave poor results above CO_2 critical temperature.

3.1.3 Calculation of wall thickness

Pipe wall thickness calculation is well established and described in national standards [12]. However, CO_2 is expected to be transported at higher pressures than natural gas and as such CO_2 pipeline will have greater wall thickness. The pipe wall thickness is obtained using **Equation 4** according to McCoy and Rubin [28] and Chandel *et al*. [12].

$$t = \frac{P_{max}D}{2(S \times F \times E - P_{\max})} \tag{4}$$

3.1.4 Pressure drop calculation

Pressure drop estimation in the dense or supercritical phase can be obtained using either liquid phase pressure drop equation or gaseous phase pressure drop equation given by **Equation 5 and 6** respectively [32].

Liquid phase pressure drop equation:

$$\Delta P_{liquid} = \frac{\rho f v^2 L}{2D_i} \tag{5}$$

Vapour phase pressure drop equation:

$$P_{out} = \sqrt{{P_{in}}^2 - Cf\rho_s LZT(\frac{{Q_s}^2}{D^5})} \tag{6}$$

At dense phase conditions, the liquid phase and vapour phase pressure drop equation has been shown to give similar results [32]. Either of the equations can therefore be used satisfactorily.

3.1.5 Pressure boosting

Pressure boosting is necessary to re-pressurise the CO_2 stream to offset pressure losses and maintain dense phase condition. This is necessary to avoid two-phase flow condition. The location of the booster station is the point where pressure is expected to drop to levels close to the two-phase boundary based on design conditions. Detailed hydraulic gradient studies are necessary to determine the number of required boosting stations. Initial compression levels could be increased to allow longer distance before boosting is needed but that will be limited by the pressure rating of the pipeline. Higher pipeline grades will mean greater CAPEX. Energy requirement for boosting depends on expected pressure level and volumetric flowrate.

3.2 Methods for estimation of capital and operating costs

Total cost of the pipeline network (CAPEX and OPEX) can be estimated through direct approach or empirical analysis. Direct approach involving cost determination at current prices is laborious and requires access to the latest prices. Empirical analysis, on the other hand, involves the use of cost estimation models. It is less laborious, quicker but less accurate. Empirical approach is adopted in this study.

Empirical cost estimation methods provides reasonable estimate of capital investment and operating cost of a CO_2 pipeline. They are crucial in early stages of pipeline design and decision making process. Hence, their accuracy is of great importance. However, limited data is usually available at the early stage of design and this has a major impact on the capability of the empirical cost correlations.

Typical empirical method applicable to onshore and offshore pipelines which gives a 'rule of thumb method' for pipeline costing has been proposed by Skovholt [17]. The method only covers estimate of initial capital expenditure and estimation accuracy is about +/- 40%. More accurate and comprehensive method was proposed by IEA GHG [27] updated in IEA GHG [30]. Other available pipeline costing models include Parker [33], McCollum and Ogden [19], Zhang *et al.* [15] and Chandel *et al.* [12]. A summary of the different pipeline cost estimation methods are presented in **Table 2**.

Table 2: CO_2 transport cost estimation studies

Study	Model	Cost Base	Outputs	Properties
Skovholt [17]	Graphic approximation	Historic project costs	Total pipeline cost	'Rule of thumb' pipeline cost estimation with +/- 40% accuracy
IEA GHG [27]	Polynomial approx. for cost estimation wit pipeline diameter and length as inputs.	Historic data from natural gas pipeline projects.	Capital cost incl. pipeline and booster pumps. Operating and maintenance costs.	Separate models for onshore and offshore pipelines, and also for different ANSI class ratings. Includes allowances for location and terrain profile Costs in year 2000 USD.
Parker [33]	Second order polynomial approx. with pipeline diameter and length as inputs.	Historic data from natural gas pipeline projects. (*Oil and Gas Journal*)	Four cost categories: materials, labour, miscellaneous, and right of way	Originally developed for Hydrogen pipelines Cost in in year 2000 USD.
IEA GHG [30]	Polynomial approx. for cost estimation.	Historic data from natural gas pipeline projects.	Capital cost incl. pipeline and booster pumps. Operating and maintenance costs.	Two separate models for Europe and North America. European model gives cost in in year 2000 EUR.
McCollum and Ogden [19]	Regression model for cost estimation. Inputs: mass flow, and length.	Other published models	Pipeline capital cost	Cost model independent of diameter. Costs estimated in year 2005 USD.
Zhang *et al.* [15]	Linear cost correlation. Inputs in cost estimate: diameter and length.	Historic data from natural gas pipeline projects. (*Oil and Gas Journal*)	Pipeline capital cost, estimate of operating and maintenance cost and annual cost.	Cost estimated in year 2007 USD.
McCoy and Rubin [28]	Double-log regression correlation for cost estimate. Inputs: length and diameter.	Historic data from US natural gas pipeline projects published between 1995 and 2005.	Four cost categories: materials, labour, miscellaneous, and right of way	Includes factors for various regions in the US. Cost is estimated in 2004 USD.
Chandel *et al.* [12]	Exponential correlation of annual cost per km and mass flow.	Other published models	Average cost per kilometre.	Cost estimated in 2008 USD.

3.3 Case Study

3.3.1 CO_2 Emitters in the Humber Region

The study area marked in **Figure 7** has big CO_2 emitters over a reasonably small geographical region [34]. CO_2 released by the main emitters are given in **Table 3**. The region is bounded by Leeds to the northwest, Rotherham in the southwest corner, and coastline at the east side border and presently has two pipeline terminals that connect onshore and offshore natural gas lines.

Emitters are classified in three tiers, namely Tier 0, 1 and 2 in terms of quantity of CO_2 released. Tier 0 emitters which include power stations, refineries and steel works release more than 1Mt CO_2 per year and are the most likely candidates for CCS. This study will therefore consider only Tier 0 emitters. They comprise twelve (12) large emitters (**Table 3**).

3.3.2 Problem Set up and Assumptions

Expected total emissions from the region with planned developments such as the Hatfield IGCC power station is over 70Mt of CO_2/year. Over 90% of these emissions come from Tier 0 large emitters [34]. Deployment of CO_2 gathering pipelines are aimed at connecting these emitters to a central onshore pipeline (**Figure 8**).

On the basis of tight emission regulatory requirements and high cost of CO_2 emission under the EU Emissions Trading Scheme, we consider that all existing Tier 0 and 1 emitters will be connected to the network by the year 2030.

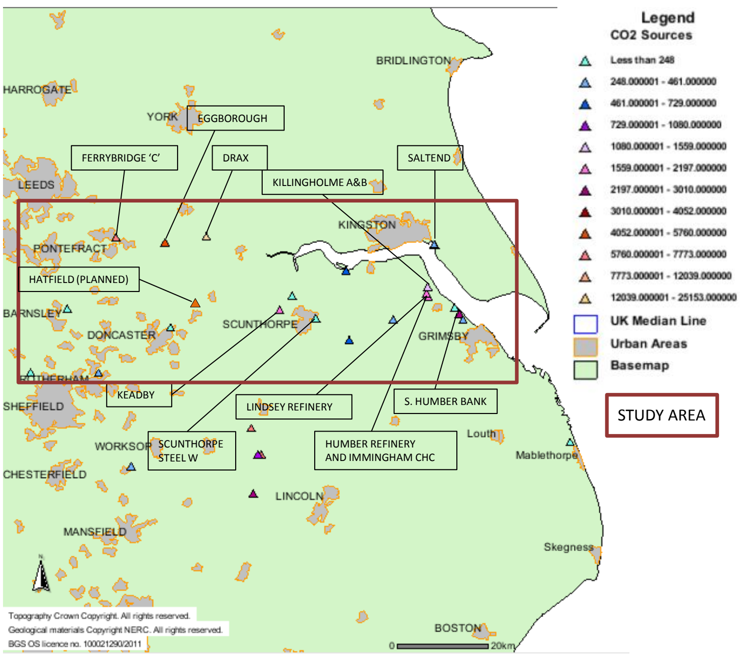

Figure 7: Study area and large CO_2 emitters in Humberside region [35]

Table 3: Emissions from Tier '0' CO_2 sources in Humber region in $ktCO_2$/year*

CO_2 Source	Location	British Geological Survey	Yorkshire Forward Survey
Drax Power	Selby	16400	22370
Eggborough PS#	Goole	5600	7676
Scunthorpe Steel Works	Scunthorpe	6780	7586
Ferrybridge ‘C’ PS#	Knottingley	6680	6200
Hatfield (Planned)	Stainforth	n/a	4800
Saltend Cogeneration	Hull	3290	3123
South Humber Bank PS#	Stalingborough	2820	2893
Immingham CHP	Immingham	n/a	2833
Humber Refinery	Immingham	2010	2400
Killingholme PS# ‘B’	Immingham	1120	2026
Lindsey Oil Refinery	Immingham	1940	2011
Keadby PS#	Scunthorpe	1730	1655
Killingholme PS# ‘A’	Immingham	1640	1497
Total Emissions		**50010**	**67070**

*Compiled from data published by BGS (based on 2002 emissions) and Yorkshire Forward (based on DEFRA [36] data)
#Power Station

For the purpose of this study, it will be assumed that the first phase of CCS deployment in this region will be driven by Drax power station, and two neighbouring big emitters, Ferrybridge ‘C’ and Eggborough power stations. It is also assumed that new Hatfield IGCC power station would connect into this network from the start of its operation. Assessment will assume that those four sources would start capturing CO_2 and transporting it through the pipeline as of year 2015. Further assessment will be carried out under the assumption that by the year 2025 all Tier 0 emitters are connected to the network. One exemption would be Saltend power station. Due to its proximity to Easington gas terminal and due to need to cross the estuary in order to join pipeline connecting other emitters, it is assumed that Saltend would have a standalone pipeline to Easington terminal. For that reason, it is not going to be included in the assessment. Pipeline connecting all other sources will end at Theddelthorpe gas terminal, which is to the north of Mablethorpe (**Figure 8 and 9**). Two options for development of CO_2 pipeline gathering network are given in **Figure 8 and 9**. In both options, pipelines terminate at Theddlethorpe gas terminal where CO_2 will be transferred into offshore pipelines.

Option 1 is based on the assumption that emissions from all sources would be developed in one main pipeline. That would mean designing the pipeline for capacity of all emitters from the very beginning of its operation although some would join into the network, as far as ten years later. In option 1, possibility of installation of a pipeline with constant diameter, and with truncated diameters is to be assessed. Pipeline routes proposed in options 1 and 2 are only indicative routes used for initial assessment purposes.

Option 2 (**Figure 9**) proposes two independent gathering lines to be developed ten years apart in order with phased roll out of CCS in mind. Phase 1 would be constructed initially to accommodate four big emitters at the north-west of the study area while phase 2 would accommodate Keadby power station, Scunthorpe steel mill, Immingham area sources (Humber and Lindsey refineries and Killingholme 'A' and 'B' power stations) and South Humber Bank power station.

Other assumptions made include:

- Design life of pipeline installations is to be 35 years for option 1 and for phase 1 of option 2 and 25 years for phase 2 of option 2.

- Amount of CO_2 produced will remain around the same level over the next 40 years. Even with growth of energy demand, new cleaner technologies will allow for more power generation at the same level of emissions.
- Capture level is assumed to be 90%. Pipeline is to be designed for maximum capture case.
- Cost of the initial CO_2 conditioning and compression is not taken into account since it will be required in all options at the same cost.
- Offshore pipeline are excluded in this study. All data used in the study are available in the public domain. The assessment covers projections up to year 2050.
- Pipeline operating pressure is to be 130bar. Minimum pressure is to be limited to 100bar. All sources connected into pipeline deliver CO_2 at the pressure of 130bar.

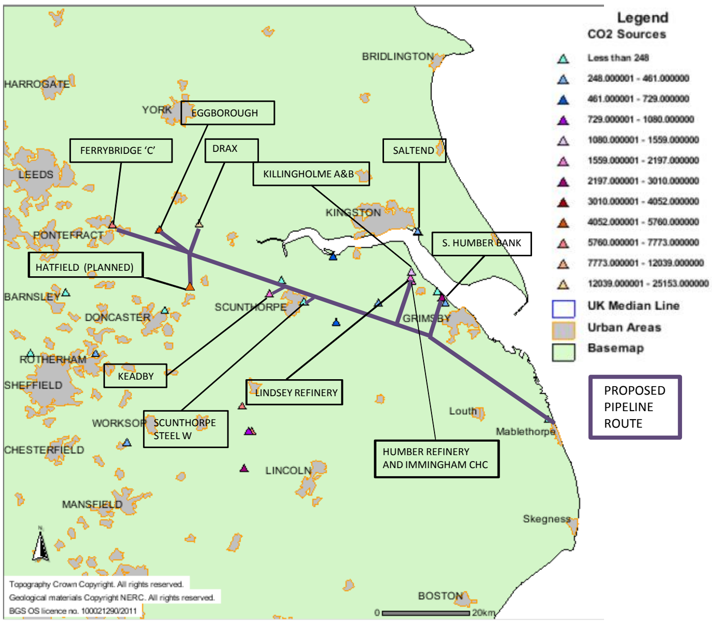

Figure 8: Pipeline outline - Option 1

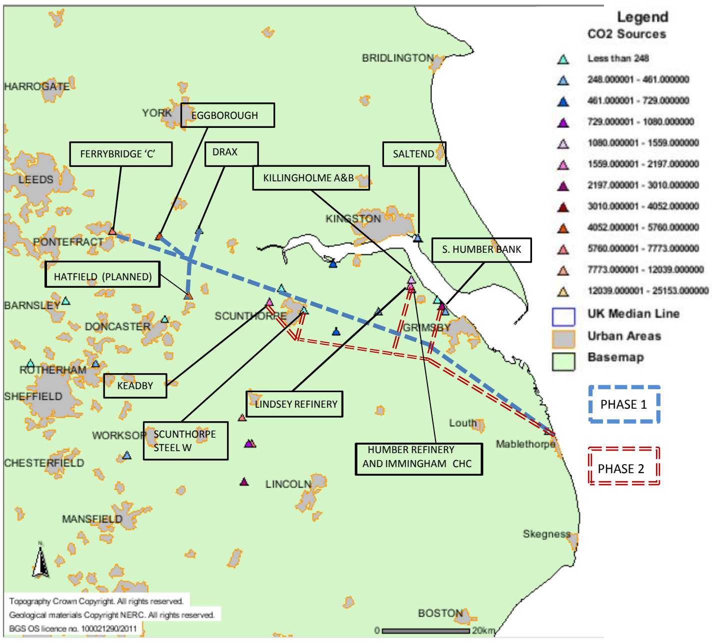

Figure 9: Pipeline outline - Option 2

- Pipeline is to be buried at 1m depth with constant temperature of the surroundings. Temperature of the stream is taken as 27°C. It is assumed that there is no heat exchange in the transport process and that temperature is constant.
- Following the analysis by Vandeginste and Piessens [20], pressure losses due to friction is considered dominant compared to losses due to bends.
- Amount of impurities in the stream are assumed to be negligible and without effect on the flow characteristics of CO_2 gas.
- Detailed topographical model of the terrain is unavailable to the authors. As a result, the terrain is assumed to be flat. Pipeline profile is consequently taken as flat without elevation differences.
- Cost of pipeline branches between the source and the main trunk line is not considered. This is because they are assumed to be the same in all options.
- Pipe material is assumed to be steel (API 5L X70).

- Selected pipe class is seamless. Seam joint factor (E) is therefore 1.0 [37].
- Class location of 1.0 is assumed. This means that the class location unit (CLU) has ten residential buildings or less. The CLU is an area extending about 201.2m on either side of the centre line of a 1.6km section of the pipeline. The corresponding safety factor (F) for a class location of 1.0 is 0.72 [37].
- Distances in Option 1 are as follows:
  - Option 1: Main route = 91.7 km
  - Junction of 4 major emitters to Keadby branch: 25.3 km
  - Keadby branch to Scunthorpe branch: 7.2 km
  - Scunthorpe branch to Immingham branch: 22 km
  - Immingham branch to S. Humber Bank branch: 8.5 km
  - S. Humber Bank branch to Teddlethorpe: 28.7 km
- Distances in Option 2 are as follows:
  - Phase 1, main route: 91.7 km
  - Phase 2, Scunthorpe jct. to Immingham: 24 km
  - Immingham branch to S. Humber Bank branch: 8.5 km
  - S. Humber Bank branch to Teddlethorpe: 28.7 km
  - Phase 2 total: 61.2 km

4 RESULTS AND ANALYSIS

4.1 Diameter Calculations

At temperature 27°C and pressure 11.5 MPa (average value between maximum and minimum pressure values of 13 and 10 MPa), density of CO_2 is taken as 827 kg/m^3 [38]. Minimum yield strength of steel pipe per API 5L X70 is 483 MPa. Design factor is taken as 0.72. Pipe class is assumed to be seamless. Pipe sizes are standard with maximum diameter of 1.32 m. Pipe roughness (ε) is taken as 10^{-4}.

Option 1

Two cases are considered. In the first case, one CO_2 gathering pipeline with constant diameter is assessed. It is designed to take emissions from all sources at maximum capacity regardless of the point at which they are attached to the pipeline. The second case considers increasing diameter in steps as additional sources are added along the pipeline route to accommodate the additional capacity. **Figure 10 and 11** provide calculation of diameters and pressure based on previously stated assumptions.

Option 2

In this option a phased approach is considered with two pipelines of constant diameter constructed at different times (**Figure 12 and 13**).

4.2 Cost estimates

In all cost calculations, adjustments were made to 2010 price levels using *Chemical Engineering Plant Cost Index,* and Bank of England's exchange rates for USD, EUR, and GBP currencies valid in August 2011 (1.00USD = 0.6112GBP and 1EUR = 0.8653GBP). Pipeline capital cost estimates are given in **Table 4**.

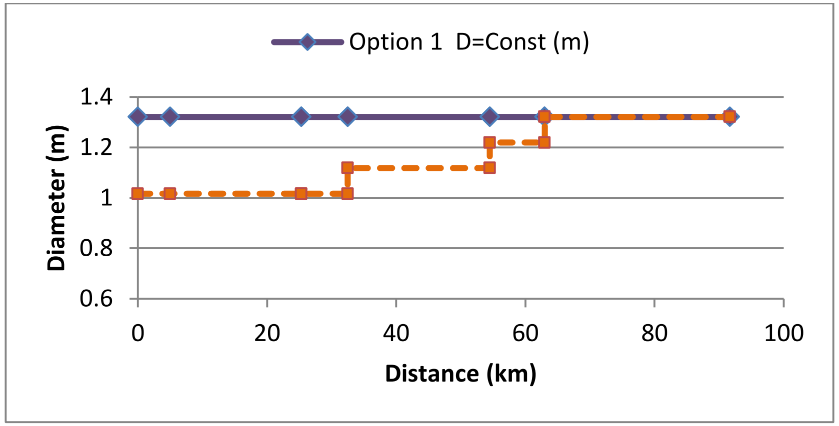

Figure 10: Option 1: Single diameter and truncated multiple diameters

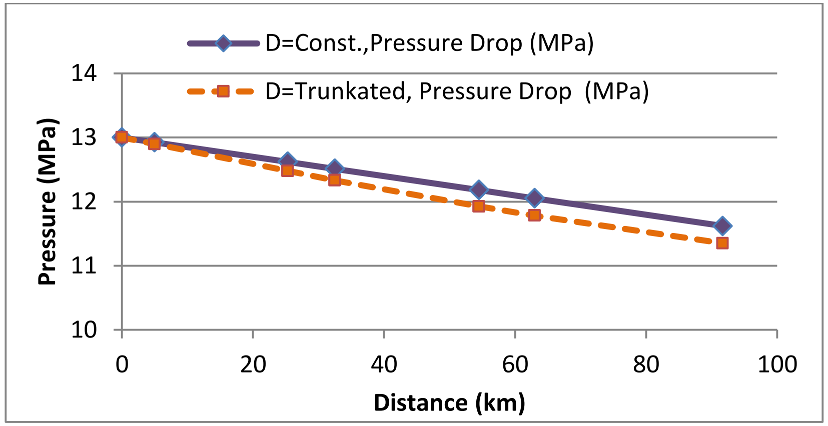

Figure 11: Option 1: Simplified pressure drop estimate along the pipeline

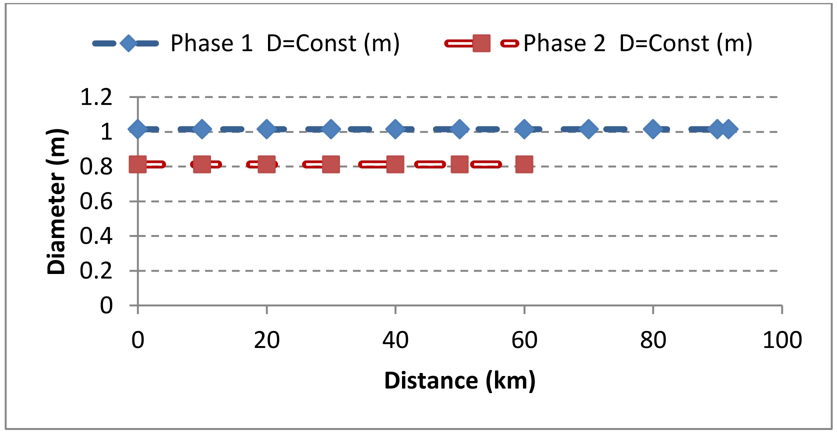

Figure 12: Option 2: Diameters for Phase 1 and Phase 2 pipelines

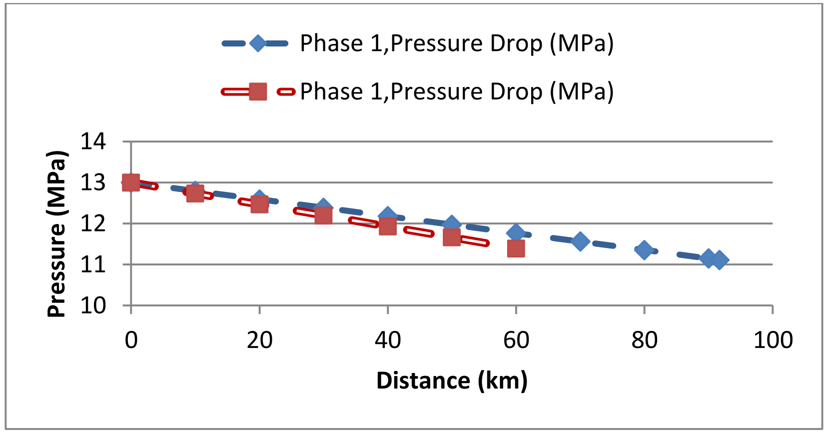

Figure 13: Option 2: Pressure drop estimate in Phase 1 & 2 pipelines

The methods that provide estimates of annual pipeline operating and maintenance costs based on pipeline length and diameter include IEA GHG [27] model, IEA GHG [30] model and Zhang *et al.* [15] model (MIT model) (**Table 5**). Results of operating and maintenance cost estimates from these models are given in **Table 5**.

Table 4: Pipeline capital cost estimates (Million GBP)

	Skovholt [17]	IEA GHG [27]	Parker [33]	IEA GHG [30]	McCollum and Ogden [19]	Zhang *et al.* [15]
Option 1, D = Const.	302.5	137.1	130.7	161.7	78	187.3
Option 1, D = Mult.	157.8	114.1	107.6	134.4	78	163.7
Option 2, Phase 1	211	85.2	87.5	100.5	66.8	144.1
Option 2, Phase 2	140.7	39.3	42.7	46.3	34.8	76.9

Table 5: Annual maintenance and operating cost estimates (Million GBP)

	IEA GHG [27]	IEA GHG [30]	Zhang *et al.* [15]
Option 1, D = Const.	1.76	4.85	0.25
Option 1, D = Mult.	1.55	4.03	0.25
Option 2, Phase 1	1.34	3.01	0.25
Option 2, Phase 2	1.00	1.39	0.16

Table 6 gives estimates of cost of CO_2 pipeline transport levelized per 1000 tonnes of transported CO_2. Annualisation of costs for models which do not provide a separate equation for this purpose was achieved using **Equation 12** given by Peters and Timmerhaus [39]. Discount rate for equipment depreciation of 10% was adopted as recommended by Peters and Timmerhaus [39].

$$C_{annual} = \frac{Capital\ cost}{((1+j)^n - 1)/j(1+j)^n} \quad (7)$$

Table 6: Levelized costs (in GBP per 1000 t of transported CO_2)

	Skovholt [17]	IEA GHG [27]	Parker [33]	IEA GHG [29]	McCollum and Ogden [19]	Zhang *et al.* [15]	Chandel *et al.* [12]
Option 1, D = Const.	545.00	264.68	235.48	300.07	140.53	338.17	325.35
Option 1, D = Mult.	284.30	223.27	193.85	249.40	140.53	294.93	325.35
Option 2, Phase 1	592.25	265.58	245.60	290.54	187.50	404.47	354.70
Option 2, Phase 2	752.02	217.24	214.80	239.91	186.00	411.02	401.19

4.3 Discussion and analysis

Pressure drop analysis in all three options assessed in the case study show that the initial pressurisation is sufficient to maintain pressure above minimum 10 MPa throughout the pipeline. Hence, booster stations are not required. Only losses arising from friction were considered ignoring elevation differences. In real conditions, this situation is unlikely because profile changes are inevitable. Closer analysis as part of detailed design process is needed to identify such points and determine if pressure boosting is required.

Pipeline capital cost estimations in **Table 4** show that the obvious choice for pipeline would be option 1 with truncated diameter. It should be noted that in order to estimate cost of truncated line, for methods which require use of diameter, each section was assessed individually and then summed up to give estimate for the whole length of the pipeline. On the other hand, methods which do not require use of diameter cannot account for difference between single and multiple diameter pipelines [19]. Estimation of pipeline capital cost for this case gives range of cost from £78 – £302.5 million for option 1 pipeline. The highest estimates are result of Skovholt [16] method and the lowest ones from McCollum and Ogden [19]. Inaccuracy of Skovholt [17] method is expected since it is developed as a rule of thumb with +/-40% accuracy. Other estimations are grouped around +/-15% of estimation resulting from IEA GHG [30] model. Cost predicted by IEA GHG [27] and Parker [33] methods are within 5% from each and about 15% lower than IEA GHG [30] while Zhang *et al.* [15] is about 15% higher than IEA GHG [30].

Estimations of operating and maintenance cost are different by a magnitude of order in case of IEA GHG [30] and Zhang *et al.* [15]. Both of these are crude estimates. In case of IEA GHG [30], annual operating and maintenance cost is taken as 3% of pipeline capital cost, while Zhang *et al.* [15] model assumes cost of 5,000 USD per mile of pipeline.

From **Table 6**, it is even more obvious that the most cost effective approach to development of CO_2 pipeline network would be as suggested in option 1, with truncated diameter. Again, estimated costs vary depending on method used; Skovholt [17] and McCollum and Ogden [19] gave the highest and lowest estimates respectively. Methods for levelized cost assessment tested in this case study, can in general be divided into two groups. One group of methods in which final cost estimation is achieved through a series of steps where costs are broken down to capital and operational expenditure, for both pipeline and/or pumping cost [27,30,33].

In the second group, a single equation is used to obtain total capital cost [12,15,19]. Only IEA GHG [27] and IEA GHG [30] methods allow for separate estimate of pressure boosting cost which in this case was not required, but in the case that it was, cost estimates from these two models would be higher, while cost estimated by other models would remain at the same level. It should also be noted that two models which are independent of diameter, and which base their cost estimation on mass flow rate (Zhang *et al.* [15] and Chandel *et al*. [12]) give close estimates of levelized cost, and in general 10-20% higher than IEA GHG [27] and IEA GHG [30].

Cost analysis carried out is in agreement with findings of McCollum and Ogden [19] and Chandel *et al.* [12] who have also analysed some of the methods presented here. It is more difficult to observe trends of cost estimates from this case study because every option presented here has either a different flow rate, length, or both flow rate and length, but a general conclusion is that capital cost is higher with higher flow rate and larger diameter, while levelized cost is lower (**Table 6**). In this case, it can be observed by comparing levelized costs of option 1 with constant diameter pipeline, and option 2 phase 1. Levelized costs calculated for this case study range from 0.14 to 0.75 GBP per tonne CO_2. This matches levelized cost predictions given in IPCC [1] on CCS for onshore pipeline 'normal' terrain conditions.

5 CONCLUSIONS AND RECOMMENDATIONS FOR FUTURE STUDY

In this paper, we evaluated methods for techno-economic assessment of newly proposed onshore CO_2 pipelines. A number of published studies in the field of CO_2 transport by pipeline have been reviewed and analysed through application on a real life case study. From the review and analysis, the properties of CO_2 such as phase behaviour and changes caused by presence of impurities in the CO_2 stream are major considerations for developing CO_2 pipeline network.

In the technical assessment, the main difficulty is presented by significant change of physical properties with relatively small changes in pressure and temperature conditions. All methods therefore involve a number of assumptions when estimating density and viscosity changes which both directly influence pressure drop and diameter calculations. These properties are usually averaged over the full length of the pipeline, or sections of it depending on the total length. The danger is that local conditions over a relatively short section can be significantly different to the rest of the pipeline. Dynamic simulation of the pipeline should be carried out using process simulation tools such as Aspen Plus® and OLGA among others to fully understand these behaviours and predict the properties more accurately. Also, actual temperature profiles can be studied using CFD tools rather than assuming constant temperatures.

The impact of gathering CO_2 captured using different capture technologies from different sources into the same pipe network was not considered in this paper. The characteristics of captured CO_2 vary depending on the composition. The composition is subject to the capture method used and the source of the CO_2. This may have a major impact on the performance of the system and should be studied and well understood earlier on. Such investigation can be conducted via whole system simulation with appropriate process simulation tools. Whole system simulations can also be helpful for understanding the impact of variations in CO_2 supply at source.

Cost assessment methods applied in the case study have provided results which are within the limits of estimations from other studies. All cost estimation methods are empirical in origin. Due to the lack of experience and data related to CO_2 pipelines cost evaluation methods are all inherently limited. It has been observed that cost estimations vary significantly depending on the method. Comparability of the methods is also limited by different input assumptions of individual methods. An adjustment to bring the cost estimates to the same year basis and currency has been carried out, but results still remained significantly spread. Flexibility of evaluation methods is also limited due to the inherent limitations of the original data set used for method development, which involve data from specific geographical regions, or limited range of pipeline lengths, diameters or flow rates.

In the future, detailed topography modelling is necessary so that actual pipeline profiles can be obtained and used to improve the results of the studies. Finally, assumption of class location of 1.0 (safety factor of 0.72) is unconservative for safe operation of CO_2 pipeline in the region. This can be improved by taking into account the actual population spread around selected channel for the onshore CO_2 pipeline. This information was not available to the author as at the time the paper was prepared.

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