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The Effect of CO_2 Purity on the Development of Pipeline Networks for Carbon Capture and Storage Schemes

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ABSTRACT

One of the key aspects relating to the transportation of anthropogenic carbon dioxide (CO_2) for climate change mitigation as part of Carbon Capture and Storage (CCS) schemes is the composition of the CO_2 stream to be transported. The specification of this stream has both technical and economic implications and, as CCS schemes start to become realised, the requirement to specify the CO_2 stream quality is becoming more important.

The aim of this work has been to analyse the effects of the composition of the CO_2 stream from post-combustion, pre-combustion and oxyfuel capture processes on the hydraulic network design and the relative costs of the network. Several key conclusions have been drawn to inform the process of specifying the CO_2 purity and to guide pipeline operators on the specification of a CO_2 stream, for dense phase pipeline operation, on the basis of hydraulic design.

The analysis has shown that impurity additions up to 2mol% did not affect the relative cost/km for the networks when compared to a pure CO_2 equivalent in terms of the pipeline internal diameter and length. However, the inlet pressure to the network is increased for all of the compositions studied and in this respect, levels of hydrogen in particular should be limited to less than 1mol% to reduce inlet pressure and thereby compression costs.

It has been demonstrated that direct connection pipelines from source to sink are the most expensive network options however, when designing a pipeline network, the size of the emitters, the phasing of entry into the network and the stability of the network in the event of interruptions in flow needs to be considered.

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1. Introduction

Carbon Capture and Storage (CCS) is recognised as one of a suite of solutions required to reduce carbon dioxide (CO_2) emissions into the atmosphere and contribute towards global climate change mitigation. In CCS schemes, CO_2 is captured from power plants or other large stationary sources and transported to appropriate geological sites either for Enhanced Oil Recovery (EOR) or for storage. Unless the capture and storage sites are co-located, all CCS schemes will involve the transportation of CO_2 from the capture plant to the storage site either via a pipeline network or by ship based transportation.

One of the barriers to the rapid implementation of CCS is the high capital cost of demonstration schemes (BBC, 2011). There is therefore an urgent requirement to reduce the costs of CCS in order that the implementation of large scale CCS schemes becomes a more viable and attractive option post demonstration (Sweeney, 2012). Studies of the costs of the full CCS chain indicate that the largest costs (whether that be in terms of the increased cost of electricity or the cost per tonne of captured CO_2) are associated with the capture process (Middleton and Bielicki, 2009; Yan et al., 2008; ZEP, 2011). However, within that chain, the cost of transportation has to be considered and cost reductions sought where possible (GCCSI, 2011).

The cost of the pipeline system has been shown to be primarily influenced by capital expenditure (CAPEX) and to be approximately proportional to the length of the network (ZEP, 2011; Knoope, 2013). Consequently, many studies have investigated the development of models to provide an optimal design for a pipeline network that minimises the present value for the capital and operating costs of the system (Brunsvold et al., 2012; Kazmierczak et al., 2009; Kuby et al., 2010; Middleton and Bielicki, 2009; Vandeginste

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Table 1
Composition ranges for CO_2 streams (Anheden et al., 2005; IEAGHG, 2011; IPCC, 2005; Kather and Kownatzki, 2011; Oosterkamp and Ramsen, 2008).

Component		Pre-combustion		Post-combustion		Oxyfuel	
		Min	Max	Min	Max	Min	Max
CO_2	vol%	95.6	99.7	99.8	99.97	85	99.94
SO_x	vol%			0.001	0.01	0.007	2.5
NO_x	vol%			0.002	0.01	0.01	0.25
H_2S	vol%	0.01	3.4				
CO	vol%	0.03	0.4	0.001	0.002		
Ar	vol%	0.03	1.3	0.003	0.045	0.01	5.7
O_2	vol%	0.03	1.3	0.003	0.03	0.01	4.7
N_2	vol%	0.03	1.3	0.021	0.17	0.01	7
H_2	vol%	0.002	1.7				
CH_4	vol%	0.035	2		0.01		
Hydrocarbons	vol%			0.003	0.01		
HCN	vol%		0.0005				
NH_3	vol%		0.003		0.005		
CH_3OH	vol%		0.02				

and Piessens, 2008). Middleton and Bielicki (2009) have indicated that there are seven parameters that should be optimised simultaneously in a comprehensive network model; the amount of CO_2 to be captured; the location of the sources; the route of the pipeline; the pipeline dimensions; the location of the sinks; the injection volume at each sink and the distribution of the CO_2 in the pipeline network. However another key parameter that needs to be considered is the purity of the CO_2.

Preliminary studies have indicated that the purity of the CO_2 has a significant impact on pipeline transport costs particularly at lower flow rates and longer distances (Yan et al., 2008) and, on this basis, it could be considered to be insufficient to consider the stream as pure CO_2 when conducting hydraulic calculations. This paper investigates the effect of the purity of the CO_2 on pipeline capacity and network development in more detail.

2. Effect Of Impurities On Pipeline Hydraulics

2.1. CO_2 Purity Specifications

Anthropogenic CO_2 captured from a power plant or any other industrial source will contain non-CO_2 components often referred to as impurities. The amount and type of impurities that could be present in the CO_2 stream from power plant capture will primarily be dependent on the capture process, the capture technology used and the fuel source. In addition, legislative and economic constraints will also play a part in determining allowable or achievable levels of certain impurities. A number of studies have been conducted into the composition of the CO_2 streams captured from power plant (Anheden et al., 2005; IEAGHG, 2011; IPCC, 2005; Kather and Kownatzki, 2011; Oosterkamp and Ramsen, 2008). The results of these studies for different capture processes and technologies are summarised in Table 1, which illustrates that a large variation exists in the published literature regarding potential levels of impurities in the CO_2 streams captured. An alternative approach to defining a CO_2 specification was taken by de Visser et al. (2008) in the Dynamis transport specification, which recommends CO_2 purity levels based on the requirements of the pipeline. It is recognised that the Dynamis specification is not a CO_2 composition and the use of the Dynamis study for defining CO_2 compositions must be considered with care (Race et al., 2012).

In the Dynamis study, safety and toxicity limits, infrastructure durability and transport efficiency are considered in the development of the specification shown in Table 2. The study presented in this paper is restricted to the effects of impurities on hydraulic behaviour, although it is highlighted that in determining a pipeline CO_2 stream specification, the effects on all aspects of pipeline design and operation must be considered (Race et al., 2012). In respect of the hydraulic behaviour it is interesting to note that, the Dynamis specification sets a limit of 4vol% on the non-condensable components of N_2, O_2, H_2, CH_4 and Ar in order to minimise the impact on pipeline capacity, capital cost and compression costs. Yan et al. (2008) have also studied the techno-economic impact of non-condensables at different levels (13%, 4% and 1% by volume) on the transportation of CO_2 from oxyfuel capture. They conclude that the limit on non-condensable components of <4vol% is a reasonable purification limit in terms of the cost balance of the CCS chain. However, they indicate that, for short distances and, where the storage conditions permit, the level of non-condensables could be raised to 10vol%, although levels as high as this may require special attention to meet regulatory requirements (*i.e.* health and safety considerations), based on the concentration of the individual impurities present.

Table 2
Pipeline specifications proposed by the Dynamis (de Visser et al., 2008).

Component	Dynamis Specification	
CO_2	vol%	>95.5
H_2O	ppm	<500
SO_x	ppm	<100
NO_x	ppm	<100
H_2S	ppm	<200
CO	ppm	<2000
N_2	vol%	Total non-condensable gases <4vol%
Ar	vol%	
O_2	vol%	
H_2	vol%	
CH_4	vol%	Aquifer <4vol% EOR <2vol%

The individual components of the capture streams considered in this paper are taken from IEAGHG (2011) and are presented in detail in Table 3. For the network study, one pre-combustion stream and one post combustion stream were chosen. These streams both have a high percentage of CO_2 and are considered to be representative of achievable levels of secondary components. Two oxyfuel compositions were also chosen, one being a relatively pure stream (oxyfuel 2) and the other containing a larger percentage of impurities (oxyfuel 1). The last ten years of development in oxyfuel combustion power plants with CO_2 capture have established that the Cryogenic Processing Unit (CPU) of the power plant could be designed to deliver CO_2 with composition ranging from as low as 80% to as high as 99.999%. It should therefore be noted that the oxyfuel 1 composition (presented in Table 1) could represent a possible low purity or worst case scenario, and the oxyfuel 2 composition could represent a possible high purity scenario. It is recognised that the stream compositions selected are not actual compositions but recommended specifications. Nevertheless, in the absence of

Table 3
Detailed composition of CO_2 streams from IEAGHG (2011).

Component		Pre-combustion	Post-combustion	Oxyfuel (1)	Oxyfuel (2)
CO_2	vol%	97.95	99.81	85.0	98.0
O_2	vol%	-	0.03	4.70	0.67
N_2	vol%	0.9	0.09[a]	5.80	0.71
Ar	vol%	0.03	-	4.47	0.59
H_2O	ppm[c]	600	600	100	100
NO_x	ppm	-	20	100	100
SO_2	ppm	-	20[b]	50	50
SO_3	ppm	-	-	20	20
CO	ppm	400	20	50	50
$H_2S + COS$	ppm	100	-	-	-
H_2	vol%	1	-	-	-
CH_4	ppm	100	-	-	-
Cricondenbar	bara	77.54	73.93	93.26	75.95

[a] Total concentration of N_2 + Ar;
[b] Total concentration of $SO_2 + SO_3$;
[c] Although the levels of water are quoted here, water is not considered in the hydraulic analysis calculations.

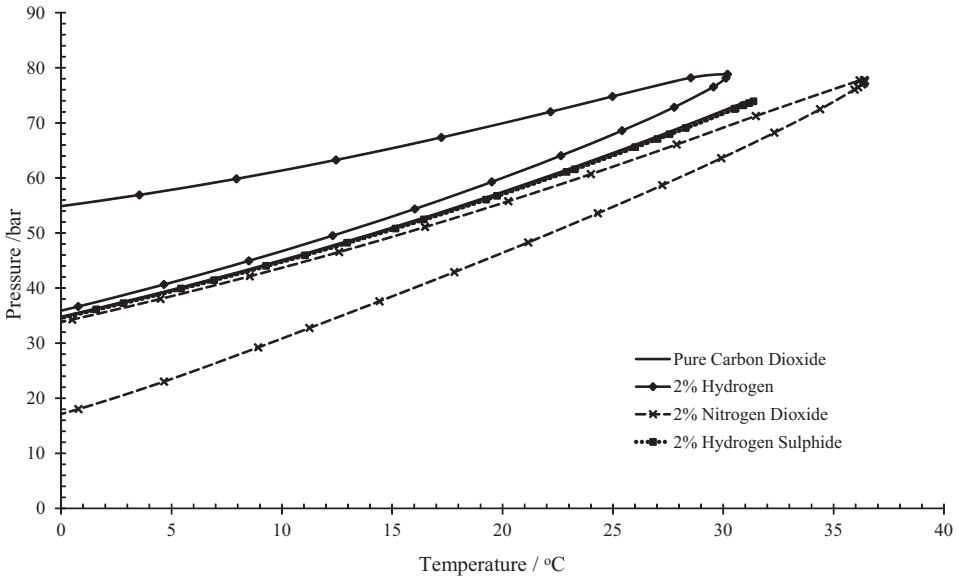

Fig. 1. Phase diagram for binary combinations of CO_2 and 2mol% H_2, H_2S and NO_2 (calculated using the Peng Robinson equation of state).

data from operational power plant capture plants, they are considered to be representative compositions for the purposes of this study.

2.2. Effects of Impurities on CO_2 Fluid Properties

The addition of impurities into the CO_2 stream affects several important parameters in the analysis of pipeline hydraulics including the phase behaviour, density, viscosity and compressibility of the fluid. Prior to the presentation of the hydraulic study, it is therefore important to understand the influence of different impurities on these CO_2 properties and ultimately on the design of the pipeline network.

2.2.1. Effect of Impurities on Phase Behaviour

To illustrate the effect of impurities on phase behaviour, consider the phase diagrams of Fig. 1 for binary combinations of CO_2 with 2mol% of hydrogen (H_2), nitrogen dioxide (NO_2) and hydrogen sulphide (H_2S). These components have been selected for illustration as they represent components that could be present in the CO_2 stream (Table 1) and they also demonstrate important behaviour in this context. All components whose critical temperature and pressure is above that of pure CO_2 will open up a two-phase region below that of pure CO_2 (e.g. H_2S and NO_2). Conversely, components with critical temperatures and pressures below those of pure CO_2 will open up a two-phase region above that of pure CO_2. The effect of each of the components considered on the phase envelope is illustrated in Table 4. Although all of the impurities raised the critical pressure, at the levels studied, components with a critical temperature below that of pure CO_2 lowered the critical temperature of the mixture relative to pure CO_2, whilst those with a higher critical temperature than pure CO_2 raised the critical temperature of the mixture.

These effects are important as the change in phase behaviour limits the allowable operating region of the pipeline. CO_2 is transported most efficiently by pipeline as a dense phase or supercritical liquid[1]. However it is essential for operating efficiency, and to prevent damage to components such as valves, pumps and compressors that the fluid remains in a single phase. Consequently, it is desirable to maintain the pressure in the pipeline above the cricondenbar of the fluid. Additions of impurities with critical temperatures and pressures below CO_2 will therefore require higher

[1] In this paper two regions are defined above the critical pressure; the "supercritical phase" which lies above the critical temperature and the "dense phase" which lies below the critical temperature.

Table 4
Relative critical pressures of key impurities and their effect on the phase envelope.

	Molecular Weight	Critical temperature (°C)	Critical pressure (bar)	Effect on phase envelope
Hydrogen	2	-240.0	13.0	Phase envelope above CO_2
Nitrogen	28	-147.0	33.9	
Carbon monoxide	28	-140.35	35.0	
Argon	40	-122.4	48.7	
Oxygen	32	-118.6	50.4	
Methane	16	-82.8	46.0	
Carbon dioxide	44	31.0	74.1	
Hydrogen sulphide	34	100.1	89.4	Phase envelope below CO_2
Sulphur dioxide	64	157.7	78.8	
Nitrogen dioxide	46	157.9	101.0	

operating pressures to be specified resulting in increased costs for compression and pumping.

2.2.2. Effect of Impurities on Density

Fig. 2 illustrates the non-linear relationship between temperature, pressure and density for pure CO_2. In general, the density of CO_2 decreases with increasing temperature and decreasing pressure, however, the behaviour is non-linear and a sharp discontinuity in density occurs close to the Vapour-Liquid Equilibrium (VLE) line due to the phase change from the liquid to gaseous phase. In this region, small changes in temperature and pressure can have large influences on density. The addition of impurities moves the location of the discontinuity to higher pressures for components with lower critical temperatures and pressures than CO_2, and to lower pressures for components with higher critical temperatures and pressures than CO_2, as shown in Fig. 3. This behaviour, and the effect on CO_2 pipeline transportation, has been discussed in Seevam et al. (2007). However, a key conclusion that is emphasised here is that lowering the inlet temperature will increase pipeline capacity as it increases the density of the fluid. In addition, limiting the amount of components with lower critical temperatures and pressures than CO_2 will also improve pipeline capacity.

2.2.3. Effect of Impurities on Viscosity

In general, the viscosity of the fluid increases with increasing pressure and decreasing temperature. A sharp discontinuity in viscosity is observed at the VLE and, in the liquid phase, the effect of temperature on viscosity is more dominant than in the gaseous phase (Fig. 4). The impact of impurities on the viscosity of CO_2 is also illustrated in Fig. 4. In the gaseous phase, the viscosity of the fluid is not significantly affected by the addition of impurities. However, in the supercritical phase, the viscosity is dramatically affected by the addition of impurities, with an increase in viscosity over pure CO_2 observed for components with higher critical temperatures and pressures than CO_2 (e.g.NO_2) and a decrease in viscosity over pure CO_2 observed for components with lower critical temperatures and pressures than CO_2 (e.g. H_2). Decreasing the viscosity will reduce the resistance to flow of the fluid in the pipeline.

3. Modelling Methodology

In this study, the effects of impurities have been studied, firstly on a single source to sink pipeline to model the effect of individual impurities on pipeline diameter and then on three network scenarios to study the impact of the product stream composition on the system size and configuration. The hydraulic modelling methodology for both of these studies is described in the following sections.

3.1. Determination of Pipeline Diameter

One of the first stages in the design of a pipeline is to calculate the required internal diameter for the anticipated flow rate. Several simple models for determining the required pipeline diameter exist, many of which form the basis of techno-economic models for CO_2 pipeline transportation (Heddle et al., 2003; Hendriks et al., 2003; IEA, 2002, 2005a,b; McCoy and Rubin, 2007; Ogden et al.,

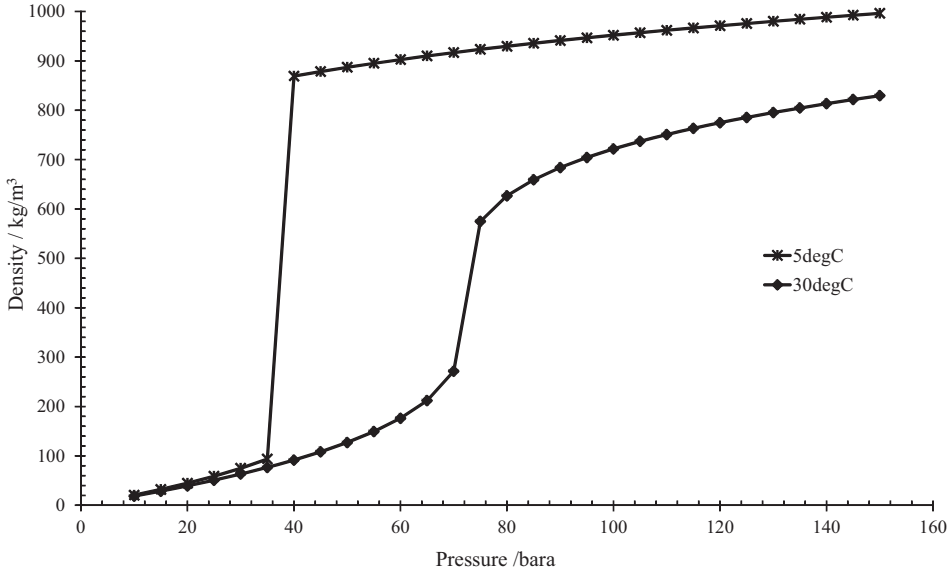

Fig. 2. Relationship between pressure, temperature and density for pure CO_2 (calculated using the Peng Robinson equation of state).

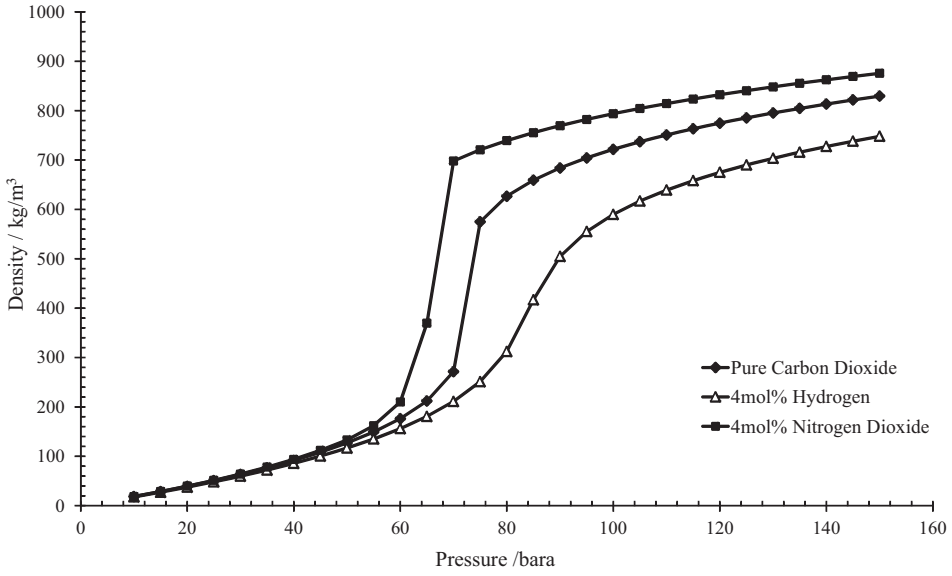

Fig. 3. Effect of impurities on the density of CO_2 for binary combinations of CO_2-4mol% H_2 and CO_2-4mol% NO_2 at 30 °C (calculated using the Peng Robinson equation of state).

2004). A review of these simple models by Ghazi and Race (2012) recommended that diameter calculations based on fluid mechanics principles (rather than mass balance equations or rules-of-thumb) should be used in techno-economic models as they require fewer initial assumptions to be made. Although these simple models are adequate for the requirements of initial pipeline sizing and costing, detailed network sizing studies require the use of more sophisticated steady-state hydraulic models which account for the effects of pressure and temperature drop along the pipeline and the consequent change in fluid properties that result.

In general, the calculation of steady state fluid flow in pipelines requires the simultaneous solution of the equations for conservation of mass, momentum and energy. From the solution of these equations, for any known fluid composition and given two of the parameters of initial pressure, final pressure or flow rate, it is possible to calculate the pressure and temperature drop along a pipeline length. Alternatively, as was conducted in this study, for a given outlet pressure, required pressure drop and flow rate, it is possible to calculate the optimum pipeline diameter. The hydraulic modelling in this study has been conducted using the PIPESIM steady-state multiphase flow simulator software (Schlumberger, 2010). The numerical procedure employed in PIPESIM is based on the method of finite differences. The pipeline is divided into segments and the pressure and temperature gradient calculations are performed in the direction of flow on each segment based on the average fluid conditions in the segment. A value of the unknown parameter is set and iteratively adjusted until the output value matches the calculated value. Once convergence has been achieved, the calculation moves to the next pipeline segment. Fig. 5 presents a flow diagram for the calculations conducted by the PIPESIM software, indicating the models that have been selected for the study.

The CO_2 physical and phase properties were calculated using the software package MultiFlash (Infochem, 2011) with the Peng-Robinson equation of state (Peng and Robinson, 1976). This

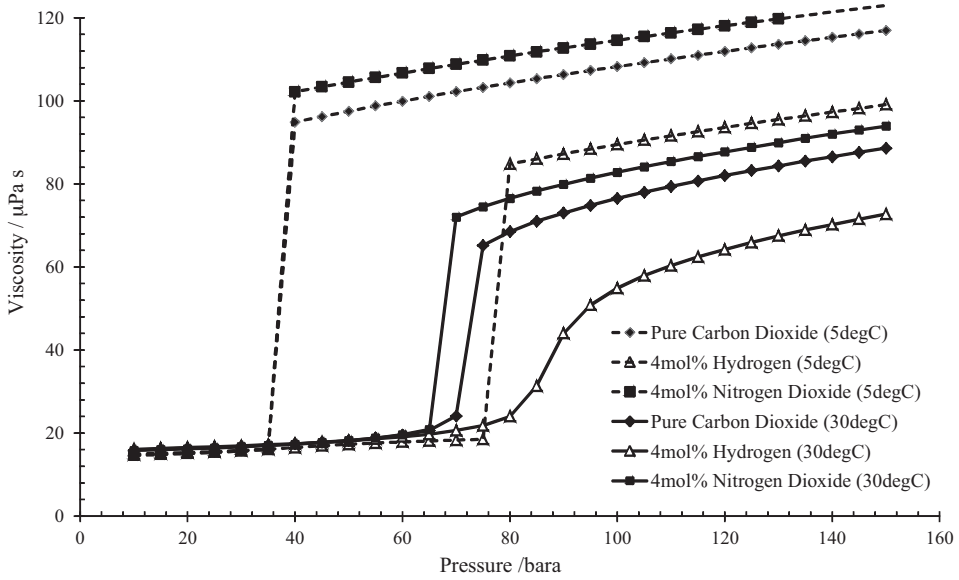

Fig. 4. Effect of impurities and temperature on the viscosity of CO_2 for binary combinations of CO_2-4mol% H_2 and CO_2-4mol% NO_2 (calculated using the Peng Robinson equation of state).

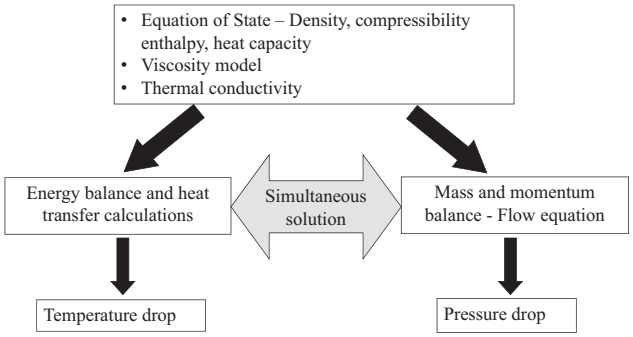

Fig. 5. Flow diagram indicating the calculations conducted in the hydraulic analysis.

equation of state was selected as it is has been shown to be sufficiently accurate for the mixture compositions and the pressure and temperature ranges explored in this paper (Li and Yan, 2009).

Although there is little published information available on the calculation of CO_2 viscosity, numerous correlations exist for calculating the viscosity in oil and reservoir fluids. In order to determine whether one of these models could be extended to calculate the viscosity of dense phase CO_2 mixtures, the calculations from two viscosity models available in PIPESIM, Pedersen (Pedersen et al., 1984) and LBC (Lohrenz et al., 1964) were compared with experimental data for pure CO_2 published by Van Der Gulik (1997). The results are presented in Fig. 6. On the basis of these results the Pedersen model was selected as it was seen to always over-predict the experimental data and therefore would be a worst-case prediction for the hydraulic calculations.

The flow equation selected for this analysis was the Beggs and Brill correlation (Beggs and Brill, 1973) with the Moody friction factor (Moody, 1944) as defined in Brill and Mukherjee (1999). The Beggs and Brill-Moody method has been demonstrated to be particularly applicable for single and multiphase fluids and has been used for the modelling of other CO_2 pipelines (Hein, 1985). This method also has the advantage that it can accurately predict small amounts of liquid formation.

The methodology adopted in PIPESIM to calculate the heat transfer coefficient between a horizontal buried pipeline and the ground surface follows the approach of Kreith and Bohn (2001) to define

Table 5
Output from emitters considered in the study AMEC (2008).

	CO_2 output per annum (Mt)	IEA tier
Station 1	0.60	1
Station 2	1.50	0
Station 3	1.65	0
Station 4	2.03	0
Station 5	2.88	0
Station 6	2.89	0
Station 7	3.12	0
Station 8	6.20	0
Station 9	7.68	0
Station 10	22.37	0

a conduction shape factor, S, from which the ground heat transfer coefficient, h_g, is calculated using the equation:

$$h_g = \frac{k_g^S}{R} \quad (1)$$

Where k_g = ground thermal conductivity and R = reference length (taken to be the pipe radius).

These models are used to calculate the heat transfer from oil and gas pipelines and it is considered that the same methodology can be applied to CO_2 pipelines as the materials and coatings used will be the same.

Once the optimum internal diameters have been calculated for each pipeline using the procedure defined above, the required external diameter and wall thickness is calculated using the thin wall formula for allowable hoop stress in PD8010-1 (2004):

$$\sigma_h = \frac{p.D_o}{20.t} \leq e.a.\sigma_{SMYS} \quad (2)$$

Where, σ_h = hoop stress (MPa), p = internal pressure (barg), D_o = external diameter (mm), t = wall thickness (mm), e = weld factor (assumed to be 1), a = design factor and σ_{SMYS} = the Specified Minimum Yield Stress (SMYS) in MPa. For the network scenarios considered, it was assumed that the pipeline would be located in a Class 1 location as defined in PD8010-1 (2004) and therefore a design factor of 0.72 was used.

In general, pipelines are supplied in standard, discrete ranges of external diameter and wall thickness. Although it is recognised that a customer can specify any external diameter and wall thickness, BS EN10208-2 (2009) indicates that, where appropriate, the

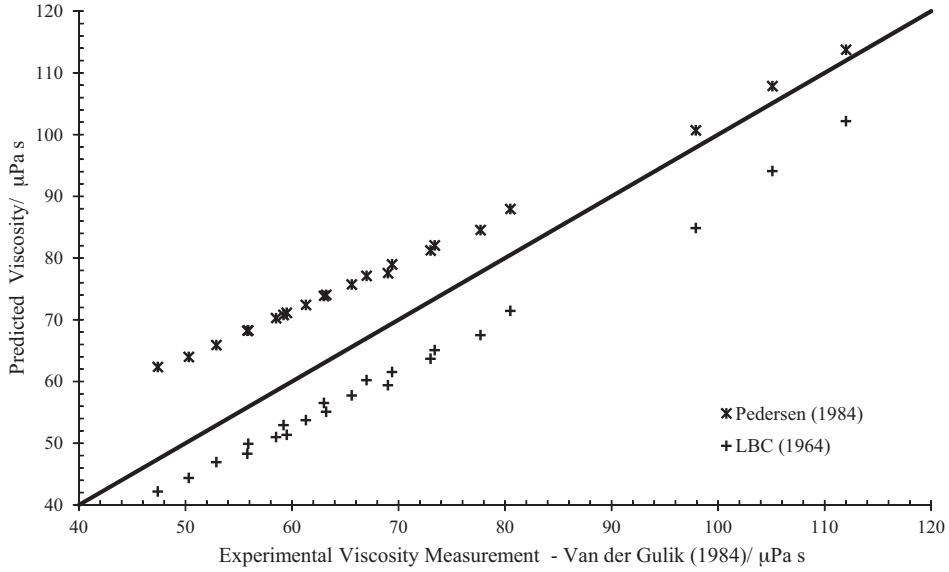

Fig. 6. Comparison of the Pedersen and LBC viscosity models with experimental data for pure CO_2.

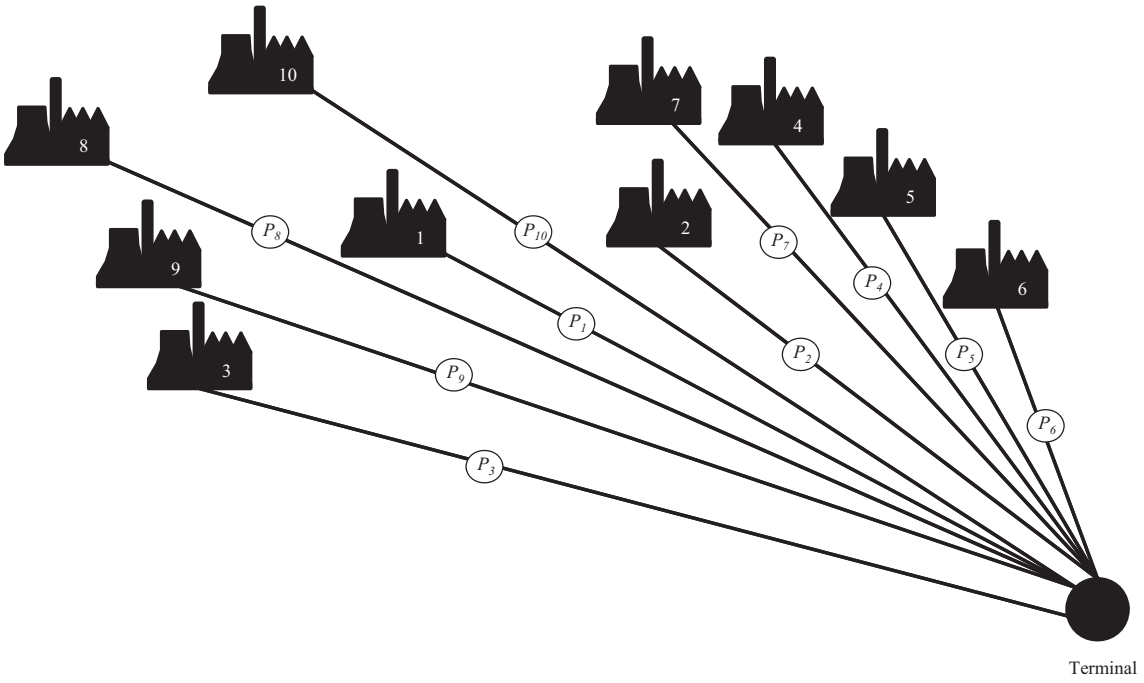

Fig. 7. Schematic representation of the Case 1 pipeline network.

external diameter and wall thickness should be within these standard ranges. Consequently, once the required external diameter and wall thickness have been determined based on the hydraulic and stress analysis constraints, the values were increased to select standard diameters and wall thicknesses as specified in BS EN10208-2 (2009).

A final check was then made with respect to the velocity of the fluid in the pipeline for the calculated flow rate and diameter. This calculation ensures that the calculated velocity is not so slow that it would affect the operation and maintenance of the pipeline and neither is it so fast that it could cause erosion of the pipeline. To check the erosional velocity the procedure outlined in API RP 14E (1991) for the calculation of erosional velocity was adopted. In order to ensure that erosion is not a threat to the pipeline, the actual velocity must be less than the erosional velocity. The erosional velocity is calculated using the following equation:

$$v_e = \frac{C}{\sqrt{\rho_{mix}}} \tag{3}$$

Where v_e = erosional velocity (m/s); ρ_{mix} = density of the fluid mixture (kg/m^3) and C = is an empirical constant as defined in API14E (1991). The value of C in Equation 3 has been determined

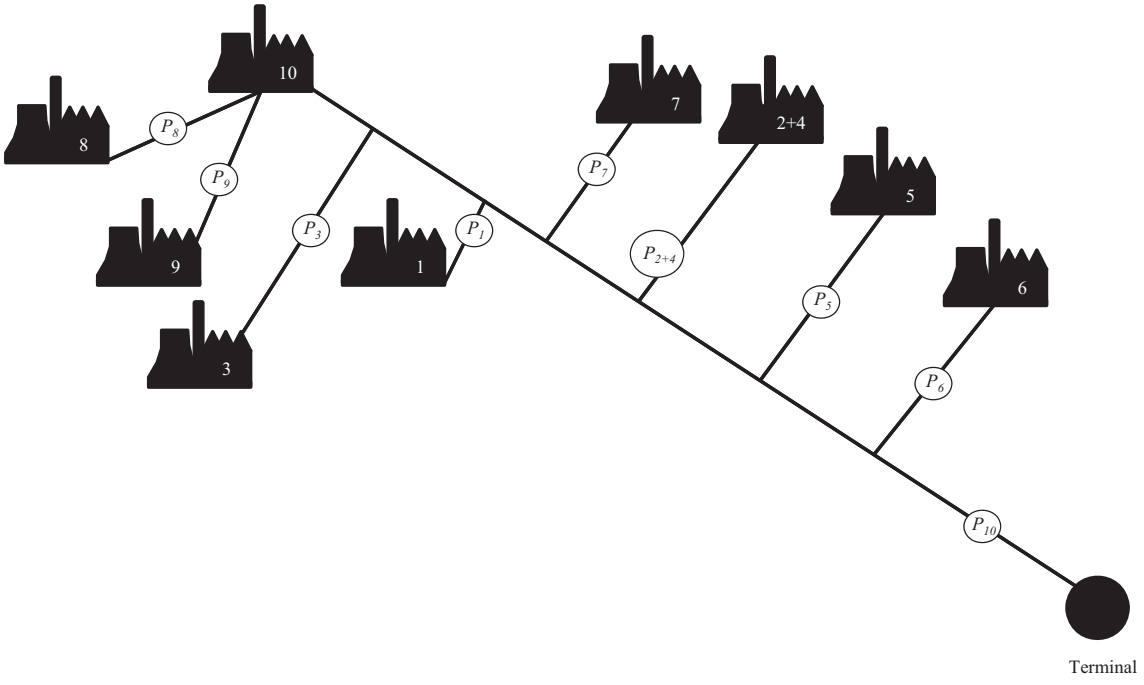

Fig. 8. Schematic representation of the Case 2 pipeline network.

Table 6
Lengths and flow rates of pipelines for network Cases 1-3.

Pipeline	Case 1 (km) Distance (km)	Pipeline Flow rate (MT/yr)		Case 2 (km) Distance (km)	Flow rate (MT/yr)	Case 3 (km) Distance (km)	Flow rate (MT/yr)
P_1	53	0.54	P_1	4	0.54	1	0.54
P_2	43	1.35	P_2 & P_4	18	3.18	17	3.18
P_3	70	1.49	P_3	7	1.49	2	1.49
P_4	43	1.83	–	–		–	
P_5	46	2.59	P_5	14	2.59	16	2.59
P_6	36	2.60	P_6	12	2.60	14	2.60
P_7	52	2.81	P_7	23	2.81	26	2.81
P_8	108	5.58	P_8	19	5.58	10	5.58
P_9	98	6.91	P_9	9	6.91	98	25.70
P_{10}	90	20.13	P_{10}	90	45.83	90	20.13
Total	639	45.83	Total	194	45.83	274	45.83

empirically from experiments conducted in the oil and gas industry and a range of values for C are provided in API RP 14E (1991) depending on the service and erosive nature of the fluid. For solids-free, continuous service a value of 122 kg/m^{2}s is recommended for C (API RP 14E, 1991) and this value has been adopted for this work. Although this value for C is considered to be conservative in the oil industry (Salama, 2000), it is recognised that there is no comparable experimental database from which to determine an appropriate value of C for CO_2 in the gaseous or dense phase.

3.2. Modelling Assumptions and Input Data

3.2.1. Fluid Conditions

As mentioned previously in Section 2.2.1, it is essential to avoid two-phase flow in the pipeline network by keeping the system pressure above the cricondenbar of the fluid for dense phase operation. Consequently, for this study, the minimum operating pressure in the system has been taken to be 10% above the cricondenbar calculated for the given fluid composition to provide an operating margin on the pipeline pressure and avoid the requirement for intermediate compression. For the study on the effect of diameter, binary combinations of CO_2 with 2mol%, 4mol% and 15mol% impurities have been considered. It is recognised that these impurity levels are not realistic or representative of potential CO_2 streams, but they have been chosen at an exaggerated level in order that the qualitative effect of each impurity can be observed. For the network study, the compositions for each of the capture technologies presented in Table 3 have been used and it has been assumed that every emitter in the network is using the same capture technology with the same composition of CO_2 for that technology.

A pressure gradient of 0.2bara/km has been assumed for the network. This pressure gradient is considered appropriate based on operating experience quoted for CO_2 pipelines in the USA (Seevam et al., 2010) and is also in line with the pressure gradients assumed in the work of Vandeginste and Piessens (2008). The inlet temperature of the flow into the pipeline was assumed to be 30 °C. Although it is noted that the output from the compressor can be as high as 40-50 °C (Farris, 1983), a lower temperature has been adopted for this study as it has been assumed that cooling would be conducted after the final stage of pressurisation in order to maximise the density of the fluid in the pipeline (as described in Section 2.2.2).

CO_2 emission data has been taken from ten power stations in a typical regional cluster (AMEC, 2008). The emitters have been classified according to the IEA Tier classification[2] (AMEC, 2008) and the annual CO_2 output considered from each emitter is presented in Table 5. In the calculations of flow rate into the pipeline, a 90% capture rate has been assumed from each emitter.

Table 7
Hydraulic model input assumptions.

Fluid Conditions	Unit	
Inlet temperature	30	°C
Pressure gradient	0.2	bar/km
Arrival pressure at terminal	Cricondenbar +10%	
Flow rate	90% of CO_2 emissions	MT/yr
Pipeline and Environmental Conditions		
Pipeline roughness	0.0457	mm
Pipeline burial depth	1.2	m
Pipeline material yield strength	450	MPa
Pipeline insulation	None	
Pipeline thermal conductivity	60.55	W/m.K
Soil thermal conductivity	2.595	W/m.K
Ground temperature	5	°C

3.2.2. Pipeline and Environmental Conditions

All of the pipelines in this study are plain carbon steel of grade EN10208 L450 (BS EN10208-2, 2009). A roughness value of 0.0457 mm has been used as the recommended value for commercial steel pipelines (Mohitpour et al., 2003). It has been assumed that the manufacture and construction standards and practices for CO_2 pipelines will be similar to those used for natural gas pipelines and therefore no insulation has been applied to the pipelines in the hydraulic model and the pipes have been buried to a depth of 1.2 m. This figure was assumed in the calculations to be representative of the maximum depth of cover required by for the construction of onshore pipelines in the UK (PD8010-1, 2004). The soil thermal conductivity has been taken to be 2.595 W/m.K, which is typical of a wet, sandy soil (McAllister, 2005). The soil thermal conductivity is considered to be constant along the whole pipeline length, although it is recognised that soil types will change over the distances modelled. The ground temperature has been taken to be 5 °C, which is the recommended design condition for natural gas pipelines in the UK (IGEM/TD/1, 2008).

3.3. Network Configurations

Three pipeline configurations have been developed to study the effect of impurities on network development;

Case 1. Direct connection between the source and the onshore terminal. This case has the largest overall length for the network at 639 km.

Case 2. A trunk line connecting Station 10 (the largest station in the network) to the terminal with the other sources feeding into this trunk line. Stations 2 and 4 share a common pipeline. This case has the shortest overall length for the network at 194 km.

[2] The IEA Tier Classification classifies CO_2 emitters by emission size: Tier 0 includes all CO_2 sources emitting over 1Mt/year, Tier 1 is made up of all sources with an output of between 50kt/year and 1Mt/year and Tier 2 includes all other sources emitting under 50kt/year.

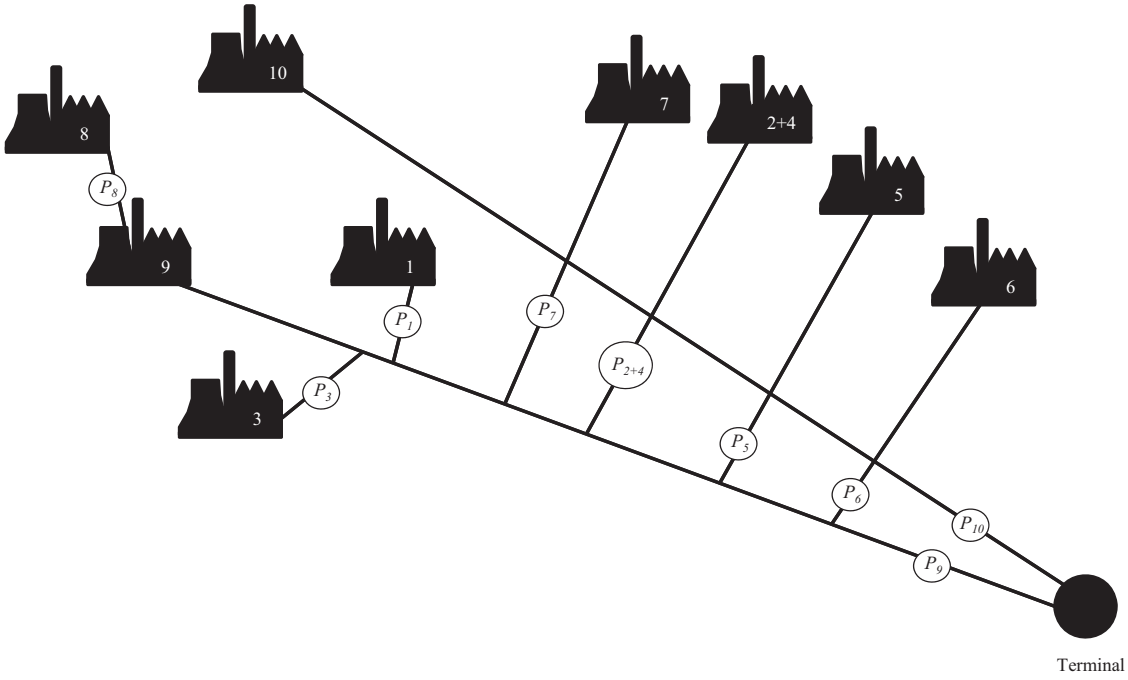

Fig. 9. Schematic representation of the Case 3 pipeline network.

Case 3. A trunk line connecting Station 9 to the terminal with the other sources (expect for station 10) feeding into this line. The CO_2 from Station 10 runs in its own directly connected pipeline from the source to the terminal. In this configuration Stations 2 and 4 share a common pipeline. Case 3 has been designed to overcome potential operational problems in Case 2 that could arise from having one large source (Station 10) connected in a pipeline network that is linking much smaller sources.

The networks are represented diagrammatically in Figs. 7–9. For Cases 2 and 3, a tree type network has been modelled with a large trunk line as this is consistent with the findings of previous studies (AMEC, 2008; Lone et al., 2010; Odenberger et al., 2008; Pershad et al., 2010). Each pipeline is defined by the label P_i, where i is the number of the power station from where the pipeline originates. The lengths of the pipelines and the total length of the network for each case are shown in Table 6. All of the pipeline connections are straight connections and no attempt has been made to account for

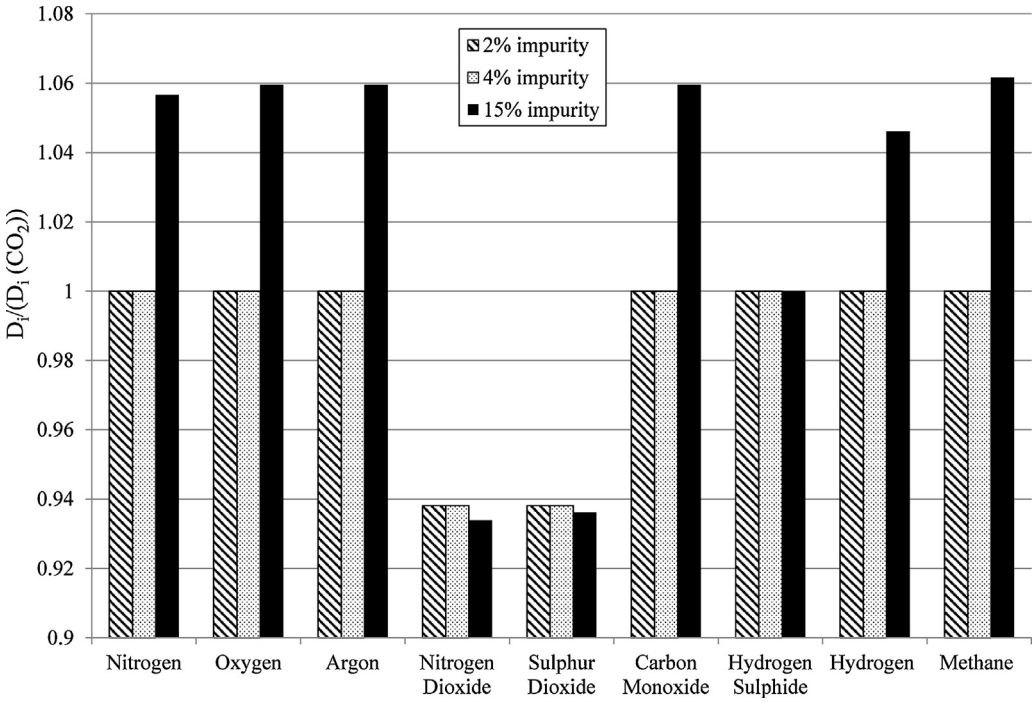

Fig. 10. Effect of impurity level on pipeline internal diameter.

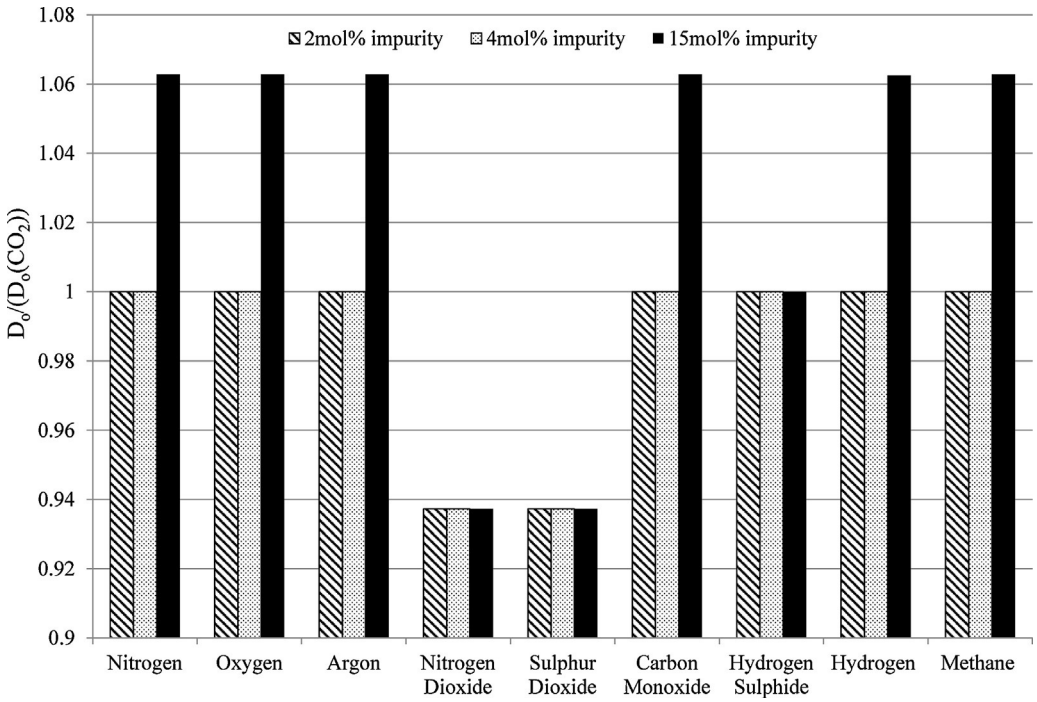

Fig. 11. Effect of impurity type on pipeline external diameter at 15mol% impurity level.

topography in the routeing. Each network case was modelled for a pure CO_2 and for each of the four capture streams presented in Table 3. The single source-to-sink pipeline that was selected was P_{10}.

3.4. Cost Modelling

The cost model that has been adopted to estimate the CAPEX of each of the pipelines and networks is that due to Ghazi and Race (2012). This model is based on the IEA model as presented in IEA (2005a,b) with the inclusion of a location factor F_L as presented in IEA (2002).

$$CAPEX(M€) = F_L.F_T.[C_1.L + C_2 + (C_3.L - C_4).]D + (C_5).L - C_6).D^2 \quad (4)$$

where, for onshore pipelines, $C_1 = 0.057$; $C_2 = 1.8663$; $C_3 = 0.00129$; $C_4 = 0$; $C_5 = 0.000486$; $C_6 = 0.000007$; D = pipeline internal diameter (inches); F_L = location factor (taken to be 1.2 for the United

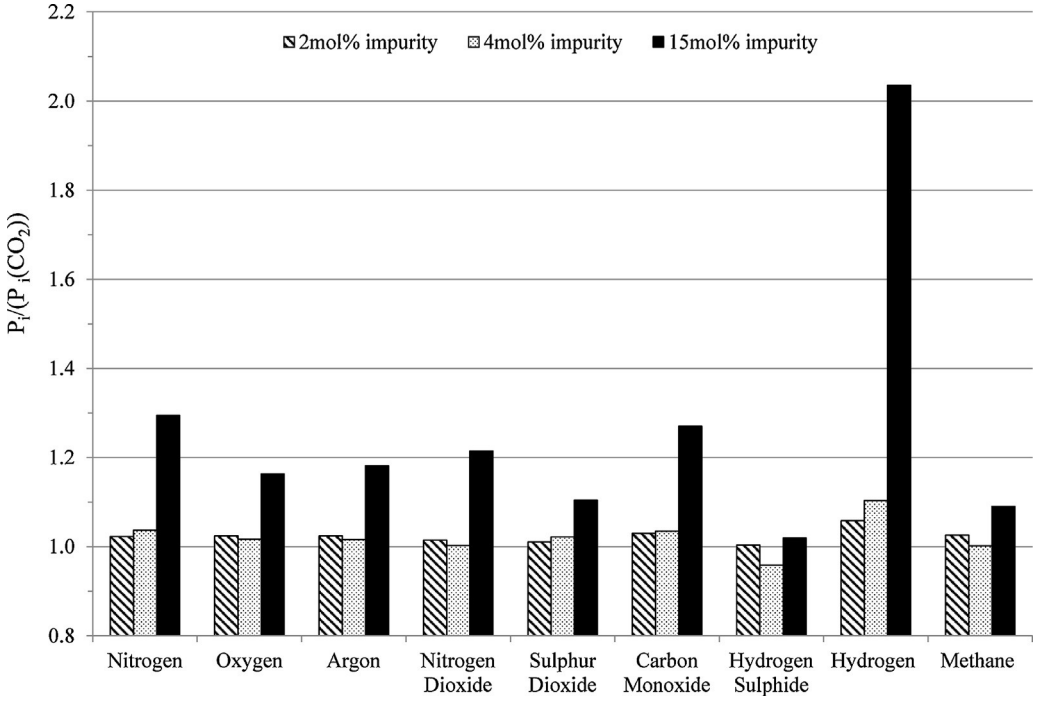

Fig. 12. Effect of impurity type on pipeline inlet pressure at 2mol%, 4mol% and 15mol% impurity level.

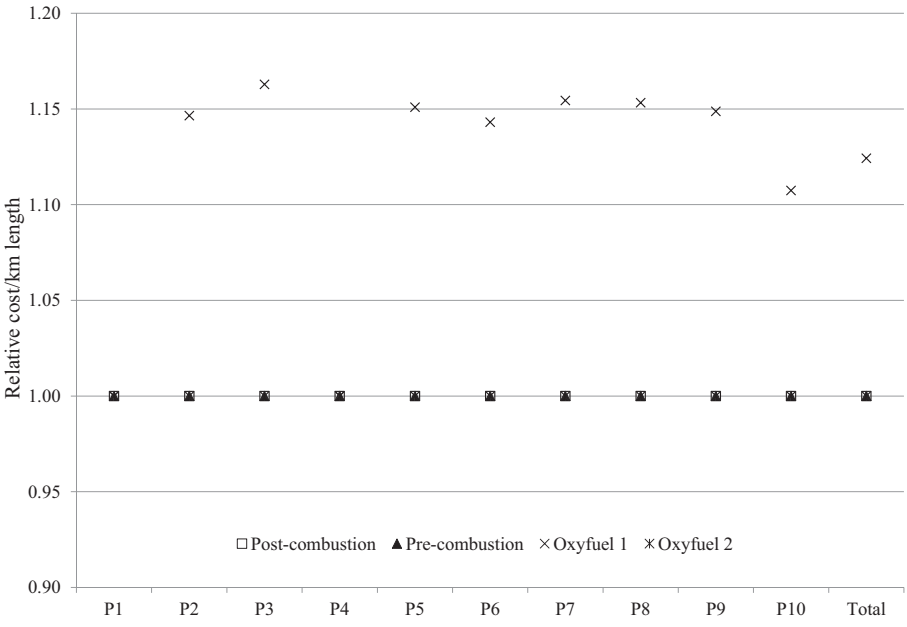

Fig. 13. Relative cost/km for each pipeline in Case 1.

Kingdom); F_T = terrain factor (taken to be 1.1 for cultivated land); L = pipeline length (km).

A summary of the input data and assumptions for the study are presented in Table 7.

4. Results

4.1. Effect of Impurities on Pipeline Diameter and Inlet Pressure

The results for the effect of impurity on pipeline internal diameter, relative to pure CO_2, are represented diagrammatically in Fig. 10. This figure indicates that, with the addition of up to 4mol% of N_2, O_2, Ar, CO, H_2, H_2S and CH_4, the impurity has no effect on the calculation of the optimum diameter size *i.e.* the internal diameter specified for the binary combination of CO_2 with impurity is exactly the same as that would be specified for pure CO_2. This result concurs with the work of de Visser et al. (2008) and Yan et al. (2008) who set a limit of 4vol% for non-condensables in terms of hydraulic efficiency. For some impurities (NO_2 and SO_2) the addition of the impurity has reduced the diameter pipeline that would be specified over pure CO_2 and the larger the level of these impurities, the smaller the pipeline that is required. However for the binary combinations of 15mol% impurity for N_2, O_2, Ar, CO, H_2 and CH_4, the diameter of the pipeline must be increased by between 4-6% over the diameter for a pure CO_2 pipeline to accommodate the higher level of impurities. This will obviously have a cost implication.

A comparison of internal diameter only takes into account the effect of impurities on pipeline capacity. However, it is also instructive to consider the effect of impurities on external pipeline diameter as illustrated in Fig. 11. For up to 4mol% impurities, there

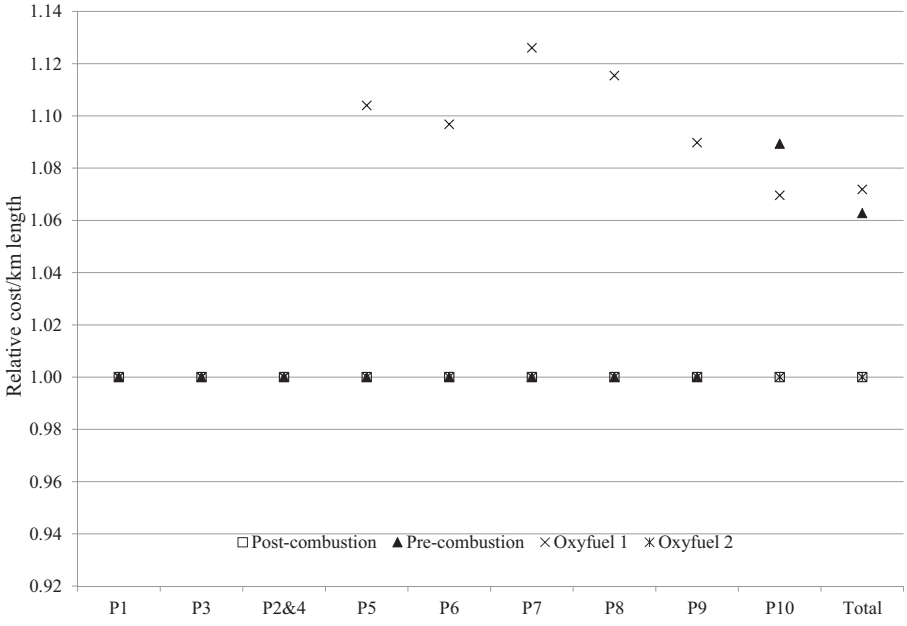

Fig. 14. Relative cost/km for each pipeline in Case 2.

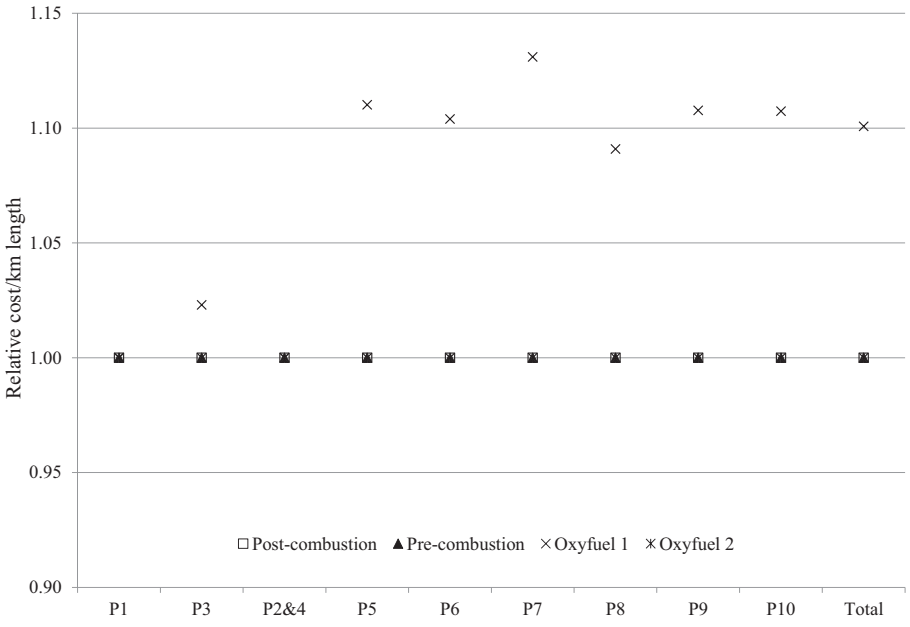

Fig. 15. Relative cost/km for each pipeline in Case 3.

is no requirement to increase the external diameter of the pipeline over that which would be specified for pure CO_2. However, for H_2 in particular, the addition of 15mol% increases the internal diameter by 4.6% over pure CO_2, but the external diameter is increased by 6.2% over pure CO_2. The reason for this is that the relative cricondenbar and therefore the inlet pressure (P_i) for the 15%H_2 mixture is significantly higher than for the other binary components (Fig. 12). Consequently, based on Equation 2, the wall thickness (and therefore the external diameter) needs to be increased for the same pipeline material. This result indicates that external rather than internal diameter should be used in cost calculations for more impure streams to take the cost of this additional material into account.

Fig. 12 also illustrates the relative effect of the types and amounts of different impurities on the inlet pressure and allows the different impurities to be ranked in terms of their efficacy in increasing inlet pressure. Therefore it can be concluded that at the 2mol% level, all impurities have a similar effect on raising the inlet pressure by 3% on average over pure CO_2. However, the addition of 15mol% H_2 doubles the inlet pressure required compared to the 2mol% mixture, whereas the addition of 15% H_2S has very little effect compared to the 2mol% mixture.

4.2. Effect of Impurities on Network Size and Configuration

4.2.1. Effect on Cost/km Length

In order to compare the costs for every pipeline in each case study, the cost/km length for each pipeline carrying the four different CO_2 streams has been calculated relative to pure CO_2 using the methodology outlined in Section 3.4. The results are presented in Figs. 13–15. From this analysis it can be seen that, for the majority

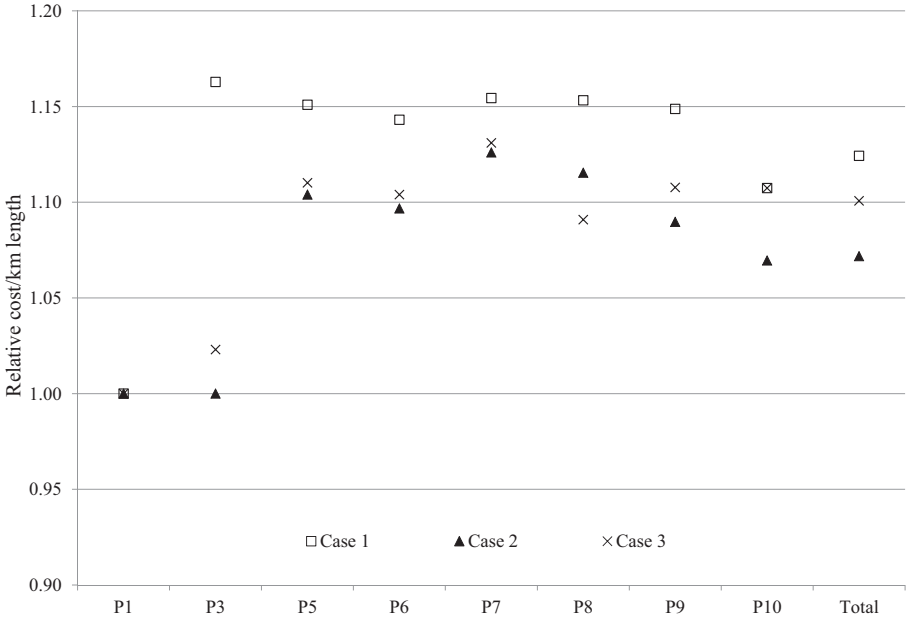

Fig. 16. Comparison of relative costs for Cases 1 to 3 using the oxyfuel 1 composition.

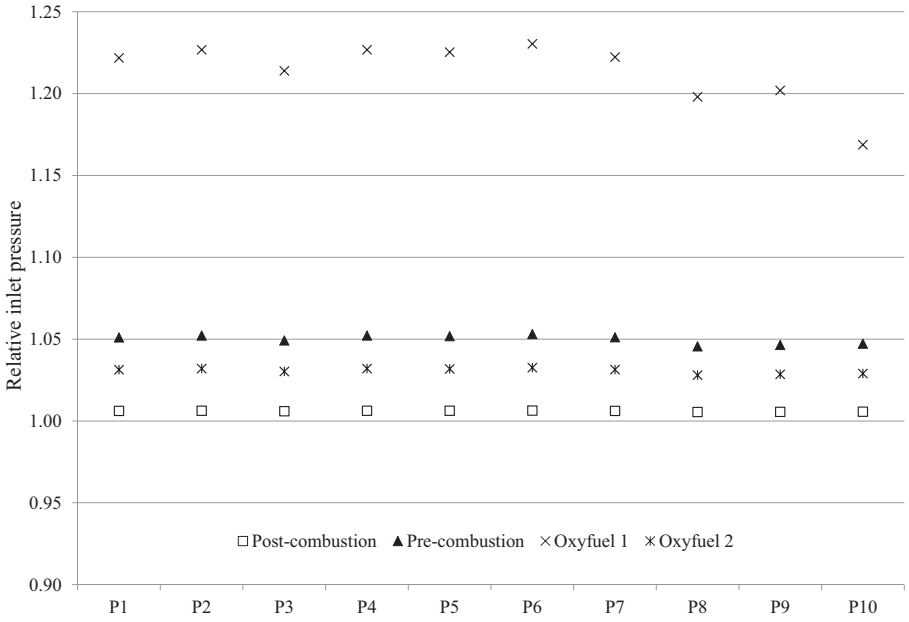

Fig. 17. Relative inlet pressure for each pipeline in Case 1.

of pipelines in Cases 1 to 3, the post-combustion, pre-combustion and oxyfuel 2 compositions have no effect on the relative cost per km for the pipelines in the network. However, the oxyfuel 1 composition, the most impure composition, can affect the relative cost of the pipeline by up to 16% and is the most expensive composition to be transported for the majority of pipelines in the three cases.

One result of note from this analysis is that for P_{10}, the main trunk line in Case 2. In this case, shown in Fig. 14, the pre-combustion composition has a higher cost/km than the oxyfuel 1 composition. The reason for this is that, although the outside diameters are the same, the internal diameters differ, highlighting again that a cost model built on external rather than internal diameter would be more appropriate.

For the oxyfuel 1 composition, a comparison has been made of the cost of the three different network cases and the results are presented in Fig. 16. These results confirm the work of other researchers that direct connection pipelines from source to terminal are more expensive than network options. Of the two networked cases, the network with the two trunk lines (Case 3) is a more expensive option overall in terms of capital cost of the network, however, it is operationally more stable. To illustrate this effect, the Case 2 network was modelled under the condition that Station 10, the largest emitter in the network, had been shut down. Consequently, the flow rate in pipeline P_{10} was reduced by 44% and the network became unstable as the flow velocity was too low for the diameter of the pipeline. As a result, during times when the flow from Station 10 was stopped for either planned or unplanned

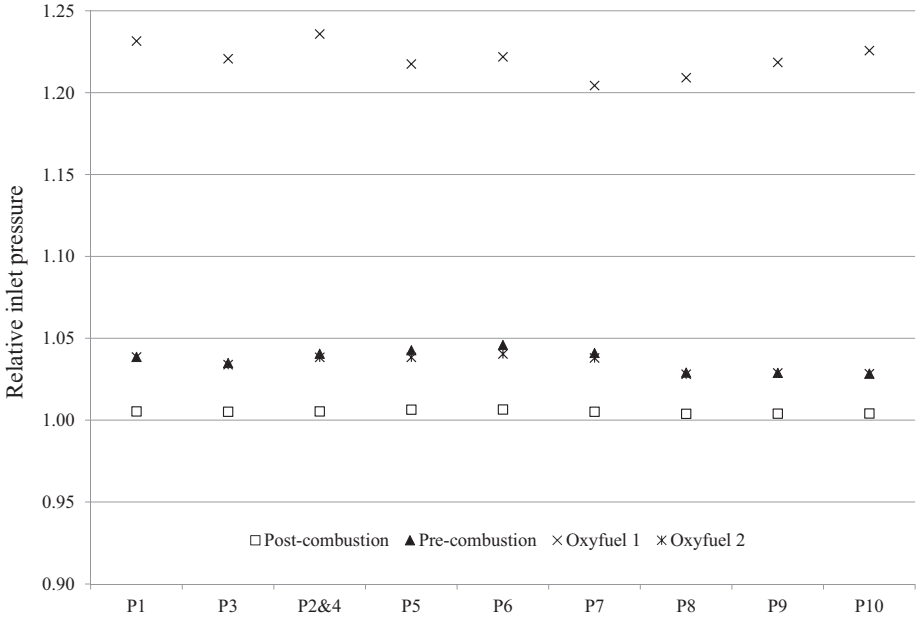

Fig. 18. Relative inlet pressure for each pipeline in Case 2.

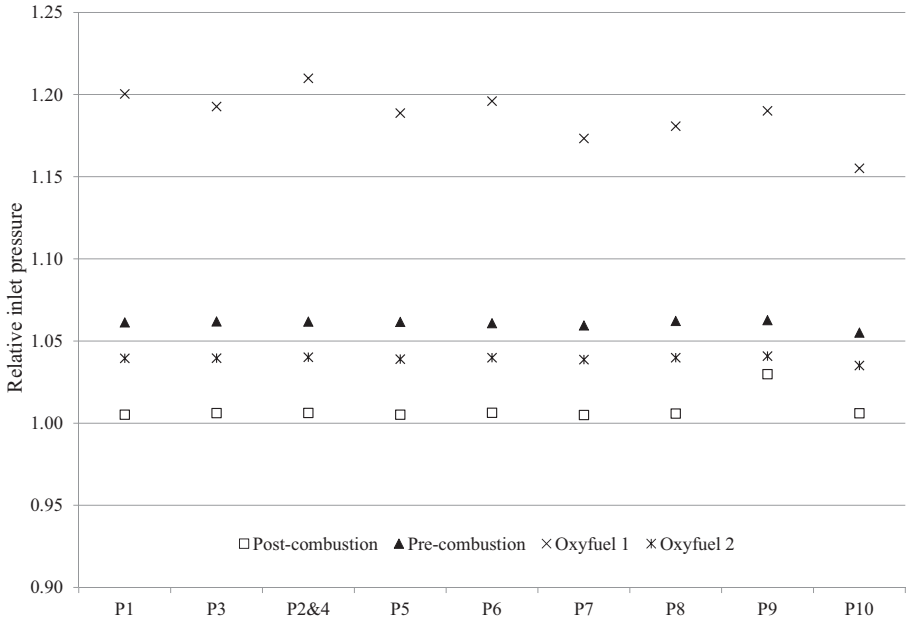

Fig. 19. Relative inlet pressure for each pipeline in Case 3.

outages, the whole pipeline network would be affected. However, in the same scenario in Case 3, the flow from Station 10 is in a single source to terminal pipeline and therefore does not dominate the flow in the network.

4.2.2. Effect on Relative Inlet Pressure

As mentioned previously, the cost calculations indicated above do not take into account the inlet pressure of the pipeline, which will affect the stress in the pipeline and consequently the wall thickness requirement. Increasing inlet pressure will also affect the cost of compression.

The relative inlet pressure (*i.e.* the inlet pressure for each pipeline relative to pure CO_2) has been calculated for each composition and each network case. The results are presented in Figs. 17–19. The results indicate that, for all cases, the oxyfuel 1 composition has the highest relative inlet pressure. This stream is the most impure stream, containing up to 15mol% impurities, and based on the analysis presented in Fig. 12, the higher the level of impurities, the higher the inlet pressure has to be to maintain the fluid in the dense phase. Similarly the post-combustion stream, which contains less than 0.2mol% impurities, has the lowest inlet pressure for all cases.

It is of particular interest to observe the results for the oxyfuel 1 and the pre-combustion compositions, which both contain 2mol% impurities. For Cases 1 and 3, the pre-combustion stream results in higher inlet pressures than the oxyfuel 2 composition. In order to account for this, comparison needs to be made between the breakdown of impurities in the stream in Table 3 and the results of Fig. 12. Although the two compositions contain 2mol% impurities, the pre-combustion case contains ~1%H_2 and ~1%N_2 which have the greatest effect in increasing the inlet pressure. The oxyfuel 2 composition contains no H_2 and the other major components of O_2 and Ar do not have as great an effect on inlet pressure. However, for Case 2, where the overall pipeline length is shorter, the effect of these differences in composition on inlet pressure is not as pronounced.

5. Conclusions

As CCS projects start to move from preliminary design to detailed design and to eventual commercial projects, the composition of the CO_2 stream will become of increasing importance in realising cost reductions in both the capture and transport parts of the chain. The aim of this work has been to analyse the effects of the composition on pipeline cost and network design for a dense phase pipeline network and several key conclusions can be drawn to inform the process of specifying the CO_2 purity on the basis of hydraulic design.

In binary combination in single pipelines, additions of up to 4mol% of impurities do not affect the diameter and wall thickness of pipeline that would be specified due to the use of discrete pipeline sizes in this analysis. In a network situation with multiple impurities present, the composition of the "non-CO_2" part of the stream does become important. The analysis presented in this paper indicates that impurity additions up to 2mol% did not affect the relative cost/km for the networks in terms of the pipeline internal diameter and length. However, the inlet pressure is increased for all of the compositions studied and this will affect the compression requirements and therefore operational cost. In this respect, it has been shown that the levels of H_2 and N_2 in particular should be limited. On the basis of the work conducted, even levels of H_2 up to 1mol% were increasing the required inlet pressure by over 6%.

An interesting conclusion can be drawn from the relative inlet pressure analysis for Case 2. In this case, the relative inlet pressures for the 98% pure streams (oxyfuel 2 and pre-combustion) were almost identical. Therefore, it has been possible to negate the effect of the H_2 in the pre-combustion stream composition by decreasing the lengths of the pipelines. It is recognised that this option might not always be possible due to other constraints of routeing caused by terrain and risk criteria but it is a choice that could be considered.

Another key observation from this work relates to the operation of networks with multiple emitters. If a number of emitters are to be incorporated into a network it is important to consider the relative contributions of each of the emitters to the flow rate. If there is one large emitter in the network and this emitter is not inputting into the system, for whatever reason, then the whole network could become unstable depending on the overall contribution that this emitter makes to the flow rate. This scenario could arise due to planned or unplanned maintenance at the source, but could also occur in the initial stages of starting up a network. It has been suggested that one scenario for infrastructure development could be to oversize trunk lines in the anticipation of future additions to the network AMEC, 2008. However, the analysis presented

in this paper indicates that, depending on the relative sizes of the emitters, this scenario would only be feasible if the largest emitters in the system were the first to input into the pipeline system.

Although this study has concentrated on the effect of impurities on the hydraulic analysis of pipelines and networks, it is reiterated that, when determining a pipeline CO_2 stream specification, the effects on all aspects of pipeline design and operation must be considered. In particular it is important to recognise that fracture control, corrosion and cracking mechanisms, hydrate formation and health and safety issues can all influence the acceptable levels of CO_2 impurities. It is further highlighted that the conclusions from this work are specific to dense phase CO_2 pipelines. Gaseous phase pipelines will have different design constraints, particularly with respect to preventing two-phase flow, where temperature is a more important constraint than pressure. Further analysis would be required to determine the impact of impurities on these constraints.

Acknowledgements

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