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Techno-Economic Models for Carbon Dioxide Compression, Transport, and Storage

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Correlations for Estimating Carbon Dioxide Density and Viscosity

David L. McCollum

Joan M. Ogden

Institute of Transportation Studies ◇ University of California, Davis

One Shields Avenue ◇ Davis, California 95616

PHONE: (530) 752-6548 ◇ FAX: (530) 752-6572

WEB: http://its.ucdavis.edu/

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University of California, Davis

One Shields Avenue

Davis, CA 95616

* Corresponding Author

E-mail: dlmccollum@ucdavis.edu

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ABSTRACT

Due to a heightened interest in technologies to mitigate global climate change, research in the field of carbon capture and storage (CCS) has attracted greater attention in recent years, with the goal of answering the many questions that still remain in this uncertain field. At the top of the list of key issues are CCS costs: costs of carbon dioxide (CO2) capture, compression, transport, storage, and so on. This research report touches upon several of these cost components. It also provides some technical models for determining the engineering and infrastructure requirements of CCS, and describes some correlations for estimating CO2 density and viscosity, both of which are often essential properties for modeling CCS. This report is actually a compilation of three separate research reports and is, therefore, divided into three separate sections. But although each could be considered as a stand-alone research report, they are, in fact, very much related to one other. Section I builds upon some of the knowledge from the latter sections, and Sections II & III can be considered as supplementary to Section I.

* Section I: *Techno-Economic Models for Carbon Dioxide Compression, Transport, and Storage* – This section provides models for estimating the engineering requirements and costs of CCS infrastructure. Some of the models have been adapted from other studies, while others have been expressly developed in this study.

* Section II: *Simple Correlations for Estimating Carbon Dioxide Density and Viscosity as a Function of Temperature and Pressure* – This section describes a set of simple correlations for estimating the density and viscosity of CO2 within the range of operating temperatures and pressures that might be encountered in CCS applications. The correlations are functions of only two input parameters—temperature and pressure—which makes them different from the more complex equation of state computer code-based correlations that sometimes require more detailed knowledge of CO2 properties and operating conditions.

* Section III: *Comparing Techno-Economic Models for Pipeline Transport of Carbon Dioxide* – This section illustrates an approach that was used to compare several recent techno-economic models for estimating CO2 pipeline sizes and costs. A common set of input assumptions was applied to all of the models so that they could be compared on an "apples-to-apples" basis. Then, by averaging the cost estimates of the models over a wide range of CO2 mass flow rates and pipeline lengths, a new CO2 pipeline capital cost model was created that is a function only of flow rate and pipeline length.

Keywords: carbon dioxide, CO2, CO_2, CCS, pipeline, transport, compression, injection, storage, sequestration, techno-economic, cost model, climate change, greenhouse gas, correlation, density, viscosity

<u>SECTION I:</u>
Techno-Economic Models for Carbon Dioxide Compression, Transport, and Storage

David L. McCollum
dlmccollum@ucdavis.edu

Institute of Transportation Studies
University of California
One Shields Avenue
Davis, CA 95616

ABSTRACT

This report provides techno-economic model equations for estimating the equipment sizes and costs of compression, pipeline transport, and injection and storage of carbon dioxide (CO_2). Models of this type are becoming increasingly important due to the recent heightened interest in carbon capture and storage (CCS) as a climate change mitigation strategy. The models described here are based on a combination of several CCS studies that have been carried out over the past few years. Because the models are laid out step-by-step, the reader should be able to understand the methodology and replicate the models on his or her own.

Keywords: carbon dioxide, CO2, CO_2, CCS, pipeline, transport, compression, injection, storage, sequestration, techno-economic, cost model, climate change, greenhouse gas

PART I: CO_2 COMPRESSION

Nomenclature

m = CO_2 mass flow rate to be transported to injection site [tonnes/day]
$P_{initial}$ = initial pressure of CO_2 directly from capture system [MPa]
P_{final} = final pressure of CO_2 for pipeline transport [MPa]
$P_{cut\text{-}off}$ = pressure at which compression switches to pumping [MPa]
N_{stage} = number of compressor stages [-]
CR = compression ratio of each stage [-]
$W_{s,\,i}$ = compression power requirement for each individual stage [kW]
Z_s = average CO_2 compressibility for each individual stage [-]
R = gas constant [kJ/kmol-K]
T_{in} = CO_2 temperature at compressor inlet [K]
M = molecular weight of CO_2 [kg/kmol]
η_{is} = isentropic efficiency of compressor [-]
k_s = (C_p/C_v) = average ratio of specific heats of CO_2 for each individual stage [-]
$W_{s\text{-}total}$ = total combined compression power requirement for all stages [kW]
$(W_s)_1$ = compression power requirement for stage 1 [kW]
$(W_s)_2$ = compression power requirement for stage 2 [kW]
$(W_s)_3$ = compression power requirement for stage 3 [kW]
$(W_s)_4$ = compression power requirement for stage 4 [kW]
$(W_s)_5$ = compression power requirement for stage 5 [kW]
N_{train} = number of parallel compressor trains [-]
W_p = pumping power requirement [kW]
ρ = density of CO_2 during pumping [kg/m^3]
η_p = efficiency of pump [-]
m_{year} = CO_2 mass flow to be transported and stored per year [tonnes/yr]
CF = capacity factor [-]
m_{train} = CO_2 mass flow rate through each compressor train [kg/s]
C_{comp} = capital cost of compressor(s) [$]
C_{pump} = capital cost of pump [$]
C_{total} = total capital cost of compressor(s) and pump [$]
C_{annual} = annualized capital cost of compressor(s) and pump [$/yr]
CRF = capital recovery factor [-/yr]
C_{lev} = levelized capital costs of compressor(s) and pump [$/tonne CO_2]
$O\&M_{annual}$ = annual O&M costs [$/yr]
$O\&M_{factor}$ = O&M cost factor [-/yr]
$O\&M_{lev}$ = levelized O&M costs [$/tonne CO_2]
E_{comp} = electric power costs of compressor [$/yr]
p_e = price of electricity [$/kWh]
E_{pump} = electric power costs of pump [$/yr]
E_{annual} = total annual electric power costs of compressor and pump [$/yr]
E_{lev} = levelized O&M costs [$/tonne CO_2]

Calculation of Compressor & Pump Power Requirements

After CO_2 is separated from the flue gases of a power plant or energy complex (i.e., captured), it must be compressed from atmospheric pressure ($P_{initial}$ = 0.1 MPa), at which point it exists as a gas, up to a pressure suitable for pipeline transport (P_{final} = 15 MPa), at which point it is in either the liquid or 'dense phase' regions, depending on its temperature. Therefore, CO_2 undergoes a phase transition somewhere between these initial and final pressures. When CO_2 is in the gas phase, a compressor is required for compression, but when CO_2 is in the liquid/dense phase, a pump can be used to boost the pressure. It can be assumed that the 'cut-off' pressure ($P_{cut\text{-}off}$) for switching from a compressor to a pump is the critical pressure of CO_2, which is 7.38 MPa. Hence, a compressor will be used from 0.1 to 7.38 MPa, and then a pump will be used from 7.38 to 15 MPa (or to whatever final pressure is desired). This line of reasoning has been adapted from [1].

$P_{initial}$ = 0.1 MPa
P_{final} = 15 MPa
$P_{cut\text{-}off}$ = 7.38 MPa

The number of compressor stages is assumed to be 5 (=N_{stage}), and the equation for the optimal compression ratio (CR) for each stage is given by Mohitpour [2]:

CR = ($P_{cut\text{-}off}$ / $P_{initial}$)^(1/N_{stage}) (where N_{stage} = 5)

The compression power requirement for each stage ($W_{s,\,i}$) is given by the following equation, which is adapted from [1] and [2].

$$W_{s,i} = \left(\frac{1000}{24*3600}\right)\left(\frac{m Z_s R T_{in}}{M \eta_{is}}\right)\left(\frac{k_s}{k_s - 1}\right)\left[(CR)^{\frac{k_s - 1}{k_s}} - 1\right]$$

Based on some assumptions and CO_2 property data from the Kinder Morgan company [3], the following values can be used in the above equation:

- For all stages:
    - R = 8.314 kJ/kmol-K
    - M = 44.01 kg/kmol
    - T_{in} = 313.15 K (i.e., 40 ^oC)
    - η_{is} = 0.75
    - 1000 = # of kilograms per tonne
    - 24 = # of hours per day
    - 3600 = # of seconds per hour
- For stage 1:
    - Z_s = 0.995
    - k_s = 1.277
    - These values correspond to a pressure range of 0.1-0.24 MPa and an average temperature of 356 K in the compressor.

- For stage 2:
  - $Z_s = 0.985$
  - $k_s = 1.286$
  - These values correspond to a pressure range of 0.24-0.56 MPa and an average temperature of 356 K in the compressor.
- For stage 3:
  - $Z_s = 0.970$
  - $k_s = 1.309$
  - These values correspond to a pressure range of 0.56-1.32 MPa and an average temperature of 356 K in the compressor.
- For stage 4:
  - $Z_s = 0.935$
  - $k_s = 1.379$
  - These values correspond to a pressure range of 1.32-3.12 MPa and an average temperature of 356 K in the compressor.
- For stage 5:
  - $Z_s = 0.845$
  - $k_s = 1.704$
  - These values correspond to a pressure range of 3.12-7.38 MPa and an average temperature of 356 K in the compressor.

Thus, the calculation for compressor power requirement must be conducted five times, since this is the number of stages that have been assumed. Although, this procedure may seem a bit more tedious than simply assuming average values for Z_s and k_s over the pressure range and using the equation only once, it is prudent to break up the calculation by stage due to the unusual behavior of CO_2's properties, which are different at each stage.

The compressor power requirements for each of the individual stages should then be added together to get the total power requirement of the compressor.

$$W_{s\text{-total}} = (W_s)_1 + (W_s)_2 + (W_s)_3 + (W_s)_4 + (W_s)_5$$

According to the IEA GHG PH4/6 report [1], the maximum size of one compressor train, based on current technology, is 40,000 kW. So if the total compression power requirement ($W_{s\text{-total}}$) is greater than 40,000 kW, then the CO_2 flow rate and total power requirement must be split into N_{train} parallel compressor trains, each operating at $100/N_{train}$ % of the flow/power. Of course, the number of parallel compressor trains must be an integer value.

$$N_{train} = \text{ROUND\_UP}\left(W_{s\text{-total}} / 40{,}000\right)$$

To calculate the pumping power requirement for boosting the CO_2 pressure from $P_{cut\text{-}off}$ (7.38 MPa) to P_{final} (15 MPa), the following equation has been adapted from [1]:

$$W_p = \left(\frac{1000 * 10}{24 * 36}\right)\left[\frac{m\left(P_{final} - P_{cut-off}\right)}{\rho \eta_p}\right]$$

(where 'm' is the CO_2 mass flow rate [tonnes/day], and the following values can be assumed: $\rho = 630\ kg/m^3$, $\eta_p = 0.75$, 1000 = # of kilograms per tonne, 24 = # of hours per day, 10 = # of bar per MPa, 36 = # of m^3*bar/hr per kW)

The following figure shows the total power requirement for the compressor(s) and pump over a range of flow rates. Notice that the dependence of compression power on flow rate, 'm', is linear, as would be expected from the equation for W_s. Also, notice how small pumping power is relative to compression power. This is because the compressor raises the CO_2 pressure from 0.1 to 7.38 MPa—a total compression ratio of 73.8—whereas the pump raises the pressure from 7.38 to 15 MPa—a total compression ratio of only 2.0.

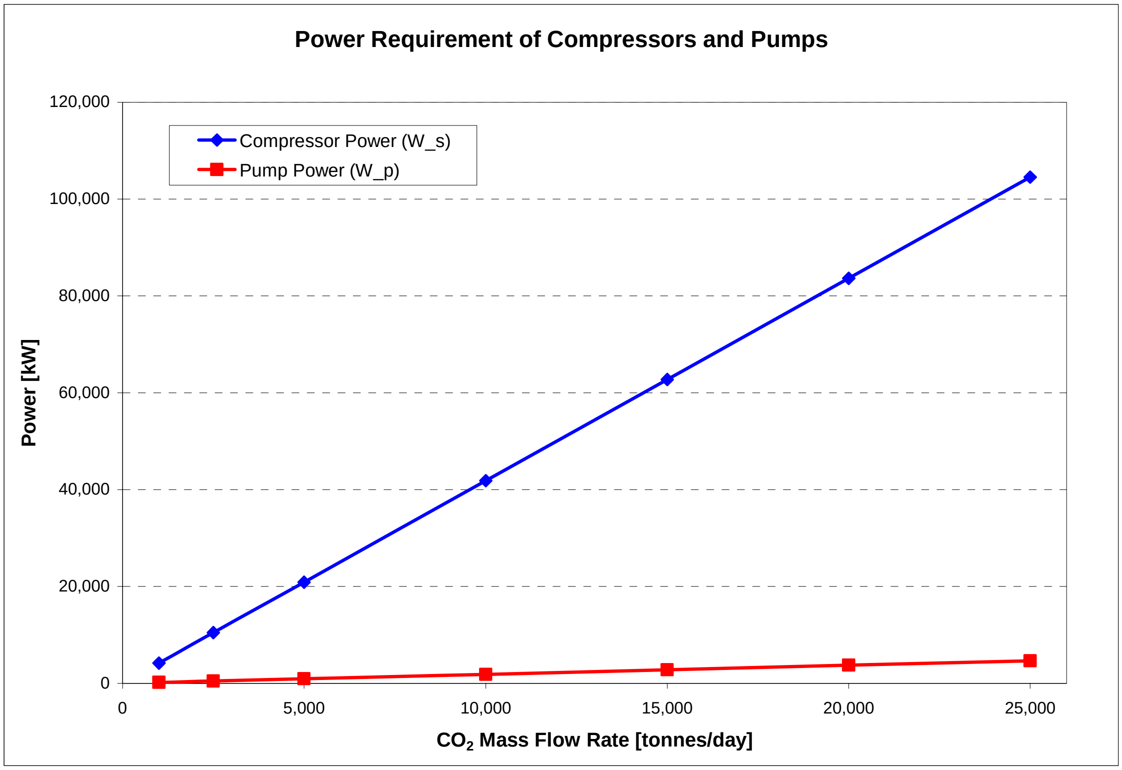


Figure 1: Power Requirement of Compressors and Pumps as a Function of CO_2 Mass Flow Rate

<u>Capital, O&M, and Levelized Costs of CO_2 Compression/Pumping</u>

*** All costs are expressed in year 2005 US$

The CO_2 mass flow rate through each compressor train (m_{train}) in units of 'kg/s' is given by:

$$m_{train} = (1000 * m) / (24 * 3600 * N_{train})$$

The capital cost of the compressor can then be calculated based on the following equation, which has been slightly adapted from Hendriks [4] and scaled up into year 2005$.

$$C_{comp} = m_{train}\ N_{train}\left[\left(0.13\times10^{6}\right)\left(m_{train}\right)^{-0.71} + \left(1.40\times10^{6}\right)\left(m_{train}\right)^{-0.60}\ln\left(\frac{P_{cut-off}}{P_{initial}}\right)\right]$$

The units on the constant terms (0.13×10^6 and 1.40×10^6) are '$/(kg/s)'. Therefore, the compressor capital cost (C_{comp}) is given in '$'.

The capital cost of the pump can be calculated based on the following equation, which has been slightly adapted from [1] and scaled up into year 2005$.

$C_{pump} = \{(1.11 \times 10^6) * (W_p / 1000)\} + 0.07 \times 10^6$

The following graph shows the capital costs of both the compressors and pumps in term of [$/kW]. As one would expect, at the higher CO_2 mass flow rates the values fall in the $1000-2000/kW range, which is consistent with other studies. Note the cost curve for compressors is not entirely smooth. This has something to do with the fact that at a certain level of compression power demand, another compressor train is added, which adds to the capital costs, but not significantly to the power demand. No such restriction is placed on pumps, so the cost curve for pumps is smooth.

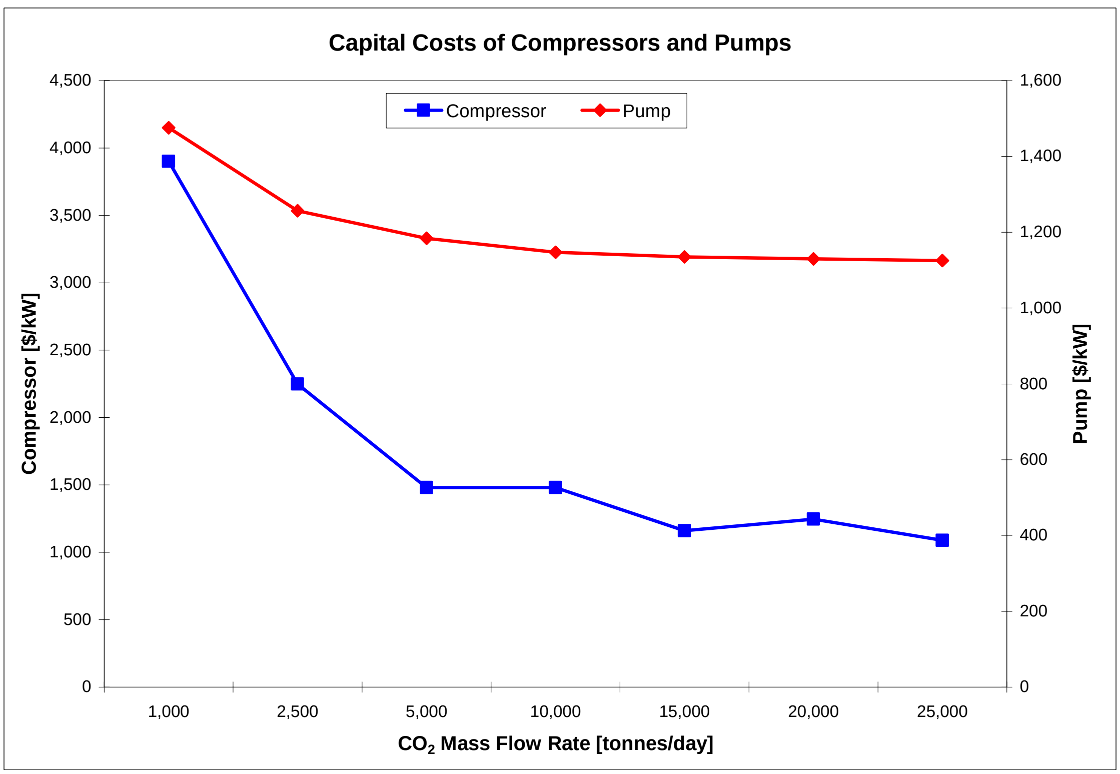


Figure 2: Capital Costs of Compressors and Pumps as a Function of CO_2 Mass Flow Rate

The total capital costs are thus:

$C_{total} = C_{comp} + C_{pump}$

The capital cost can be annualized by applying a capital recovery factor (CRF) of 0.15.

$C_{annual} = C_{total} * CRF$ (where CRF = 0.15/yr)

The total amount of CO_2 that must be compressed every year is found by applying a capacity factor (CF) of 0.80.

$m_{year} = m * 365 * CF$ (where CF = 0.80)

The levelized capital costs (C_{lev}) are thus:

$C_{lev} = C_{annual} / m_{year}$

The annual operation and maintenance costs ($O\&M_{annual}$) can be found by applying an O&M factor ($O\&M_{factor}$) of 0.04 to the total capital cost.

$O\&M_{annual} = C_{total} * O\&M_{factor}$ (where $O\&M_{factor} = 0.04$)

The levelized O&M costs ($O\&M_{lev}$) are thus:

$O\&M_{lev} = O\&M_{annual} / m_{year}$

The total electric power costs of the compressor (E_{comp}) and pump (E_{pump}) are calculated by multiplying the total power requirement by the capacity factor (CF) of 0.80 and price of electricity (p_e). It can be assumed that the electricity price is \$0.065/kWh, based on estimates by Kreutz et al. [17] for a coal-to-hydrogen plant that employs CO_2 capture.

$E_{annual} = E_{comp} + E_{pump} = p_e * (W_{s\text{-}total} + W_p) * (CF * 24 * 365)$
(where p_e = \$0.065/kWh, and CF = 0.80)

The levelized power costs (E_{lev}) are thus:

$E_{lev} = E_{annual} / m_{year}$

Finally, the total annual and levelized costs of CO_2 compression/pumping are:

Total Annual Cost [\$/yr] = $C_{annual} + O\&M_{annual} + E_{annual}$

Total Levelized Cost [\$/tonne CO_2] = $C_{lev} + O\&M_{lev} + E_{lev}$

The following two figures show the contribution of capital, O&M, and power to the total levelized cost of CO_2 compression/pumping. The reason for the cost curves not being smooth is because of the maximum power constraint of 40,000 kW per compressor train. In other words, as the flow rate of CO_2 increases, the compression power reaches a threshold point where a new compressor train is needed. This new compressor train causes a spike in the capital cost (and thus, O&M and total costs). The total power requirements and cost, however, are unaffected by the number of compressor trains that are required. Furthermore, the figures show that there are economies-of-scale associated with CO_2 compression/pumping—i.e., the capital cost becomes a smaller percentage of total cost as the CO_2 flow rate increases. The last figure shows the dependence of levelized power cost and, thus, total levelized cost on the price of electricity. Since electric power is so important to the process of CO_2 compression/pumping, it makes up an increasingly larger share of total costs as electricity becomes more expensive.

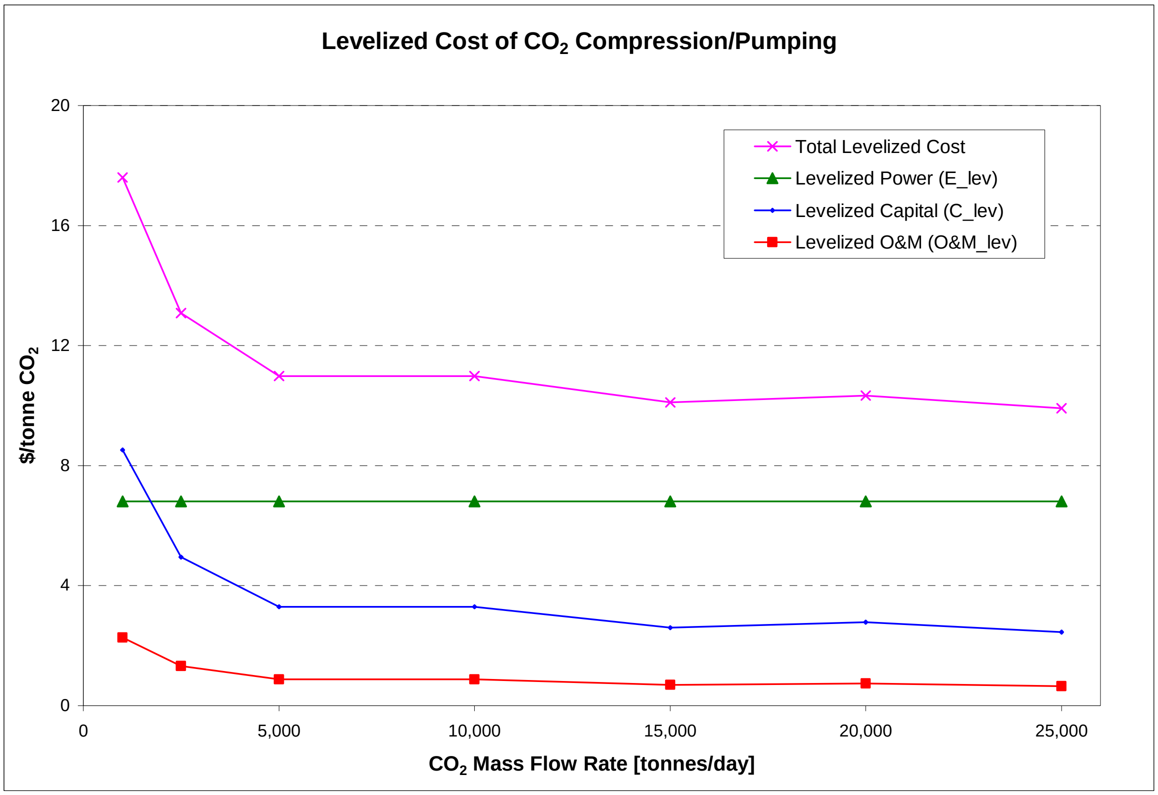


Figure 3: Levelized Cost of CO_2 Compression/Pumping as a Function of CO_2 Mass Flow Rate

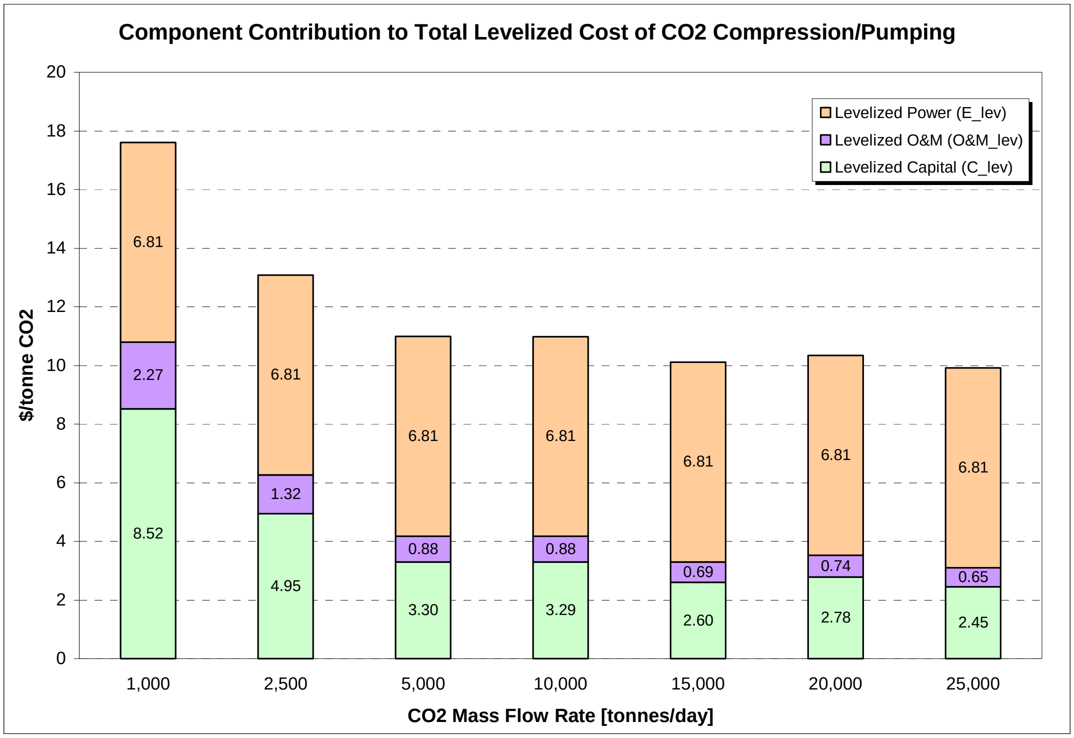


Figure 4: Contribution of Capital, O&M, and Power to Total Levelized Cost of CO_2 Compression/Pumping (Dependence on CO_2 Mass Flow Rate)

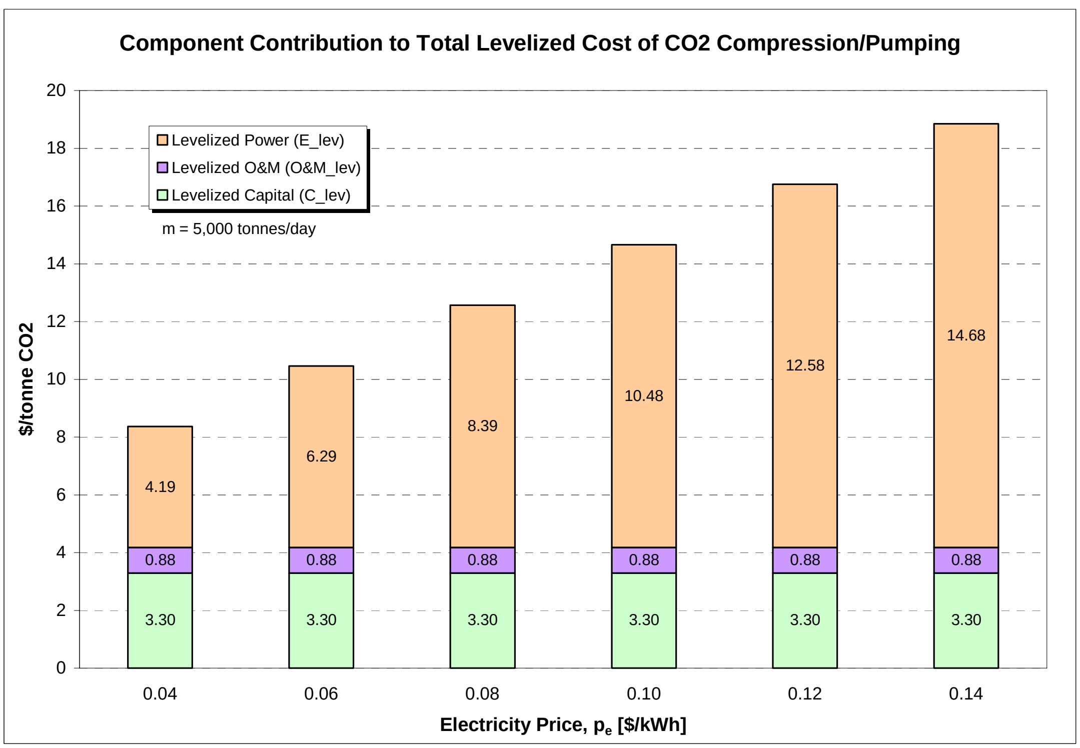


Figure 5: Contribution of Capital, O&M, and Power to Total Levelized Cost of CO_2 Compression/Pumping (Dependence on Electricity Price)

PART II: CO_2 TRANSPORT

Nomenclature

D = pipeline diameter [in]
m = CO_2 mass flow rate in pipeline [tonnes/day]
P_{in} = inlet pipeline pressure [MPa]
P_{out} = outlet pipeline pressure [MPa]
P_{inter} = intermediate pipeline pressure [MPa]
ΔP = pressure drop in pipeline = P_{in} - P_{out} [MPa]
T = CO_2 temperature in pipeline [^oC]
μ = CO_2 viscosity in pipeline [Pa-s]
ρ = CO_2 density in pipeline [kg/m^3]
ε = pipeline roughness factor [ft]
Re = Reynold's number [-]
F_f = Fanning friction factor [-]
L = pipeline length [km]
C_{cap} = pipeline capital cost [$/km]
C_{total} = total pipeline capital cost [$]
F_L = location factor [-]
F_T = terrain factor [-]
CRF = capital recovery factor [-/yr]
C_{annual} = annualized pipeline capital cost [$/yr]
$O\&M_{annual}$ = annual O&M costs [$/yr]
$O\&M_{factor}$ = O&M cost factor [-/yr]
CF = capacity factor [-]
m_{year} = CO2 mass flow delivered to injection site per year [tonnes/year]

Calculation of Pipeline Diameter

The equation for calculating pipeline capital cost (shown in the next section) is not a function of diameter. Nevertheless, when conducting a techno-economic analysis, it may be useful to estimate the diameter size for other reasons. Thus, the methodology for calculating pipeline diameter is shown here.

Since the calculation of pipeline diameter is an iterative process, one must first guess a value for diameter (D). A reasonable first approximation is D = 10 inches.

The process also requires knowledge of the CO_2 temperature (T) and pressure (P_{inter}) in the pipeline. P_{inter} is based on the pipeline inlet pressure (P_{in}, i.e. the pressure of CO_2 leaving the power plant or energy complex) and the pipeline outlet pressure (P_{out}, i.e. the pressure of CO_2 at the end of the pipeline—the injection site).

$$P_{inter} = (P_{in} + P_{out}) / 2$$

Furthermore, an estimation of the density (ρ) and viscosity (μ) of CO_2 in the pipeline (approximated at T and P_{inter}) is also required. Since CO_2 exhibits unusual trends in its properties over the range of temperatures and pressures that would be experienced in pipeline transport, it is difficult to provide just one value for either density or viscosity here. Therefore, the reader is referred to one of two CO_2 property websites, [5] and [6], or to the set of correlation equations of McCollum [7]. Each of these references provide an easy way of obtaining CO_2 density and viscosity if one knows only two basic parameters—temperature and pressure.

The Reynold's number (Re) and Fanning friction factor (F_f) for CO_2 fluid flow in the pipeline are calculated by the following equations from [8]:

$$Re = (4*1000/24/3600/0.0254)*m / (\pi*\mu*D)$$

$$F_f = \frac{1}{4\left[-1.8\,\log_{10}\left\{\frac{6.91}{\mathrm{Re}} + \left(\frac{12(\varepsilon/D)}{3.7}\right)^{1.11}\right\}\right]^2}$$

(where ε = 0.00015 ft is assumed by [8])

The pipeline diameter (D) is calculated by the following equation, which is adapted from [8]:

$$D = (1/0.0254) * [\ (32*F_f*m^2)*(1000/24/3600)^2\ /\ (\pi^2*\rho*(\Delta P/L)*10^6/1000)\ ]^{(1/5)}$$

Finally, since the process for calculating pipeline diameter is iterative, one needs to compare the calculated diameter from this last equation with the value that was initially guessed at the beginning of the process. If there is much difference between the two, then the process must be repeated over and over again until the difference between iterations is satisfactorily small.

Capital, O&M, and Levelized Costs of CO_2 Transport

*** All costs are expressed in year 2005 US$

The equations for estimating onshore pipeline capital cost are given by McCollum [9].

$$C_{cap} = 9970 * (m^{0.35}) * (L^{0.13})$$

$$C_{total} = F_L * F_T * L * C_{cap}$$

Notice that the capital cost is scaled up by a location factor (F_L) and a terrain factor (F_L). A full list of these factors is provided in [1]. A short list is reproduced here:

F_L: USA/Canada=1.0, Europe=1.0, UK=1.2, Japan=1.0, Australia=1.0.

F_T: cultivated land=1.10, grassland=1.00, wooded=1.05, jungle=1.10, stony desert=1.10, <20% mountainous=1.30, >50% mountainous=1.50

The capital cost can be annualized by applying a capital recovery factor (CRF) of 0.15.

$C_{annual} = C_{total} * CRF$ (where CRF = 0.15/yr)

The O&M costs are calculated as 2.5% of the total capital cost. This value is approximately the average O&M factor from a handful of studies on CO_2 pipeline transport [1], [8], [10], [11], [12]. To be precise, [1] and [8] do not use an O&M factor for estimating O&M costs; rather, they use a per-mile cost and an equation, respectively. Their estimates, however, are close to 2.5% of the total capital cost over the range of CO_2 flow rates and pipeline lengths considered here.

$O\&M_{annual} = C_{total} * O\&M_{factor}$ (where $O\&M_{factor} = 0.025$)

The total annual costs are thus:

Total Annual Cost [$/yr] = $C_{annual} + O\&M_{annual}$

The total amount of CO_2 that must be transported every year is found by applying a capacity factor (CF) of 0.80.

$m_{year} = m * 365 * CF$ (where CF = 0.80)

And the levelized cost of CO_2 transport is given by:

Levelized Cost [$/tonne CO_2] = (Total Annual Cost) / m_{year}

The following figures show the onshore pipeline capital cost (C_{cap}) and levelized cost, as calculated by the above equations, over a range of CO_2 mass flow rates and pipeline lengths. From these figures, it is easy to see that for capital cost there is a stronger dependence on flow rate than on length. This is to be expected since, in the equation for C_{cap}, the exponent on the flow rate term, ‘m’, is larger than the exponent on the length term, ‘L’ (0.35 vs. 0.13).

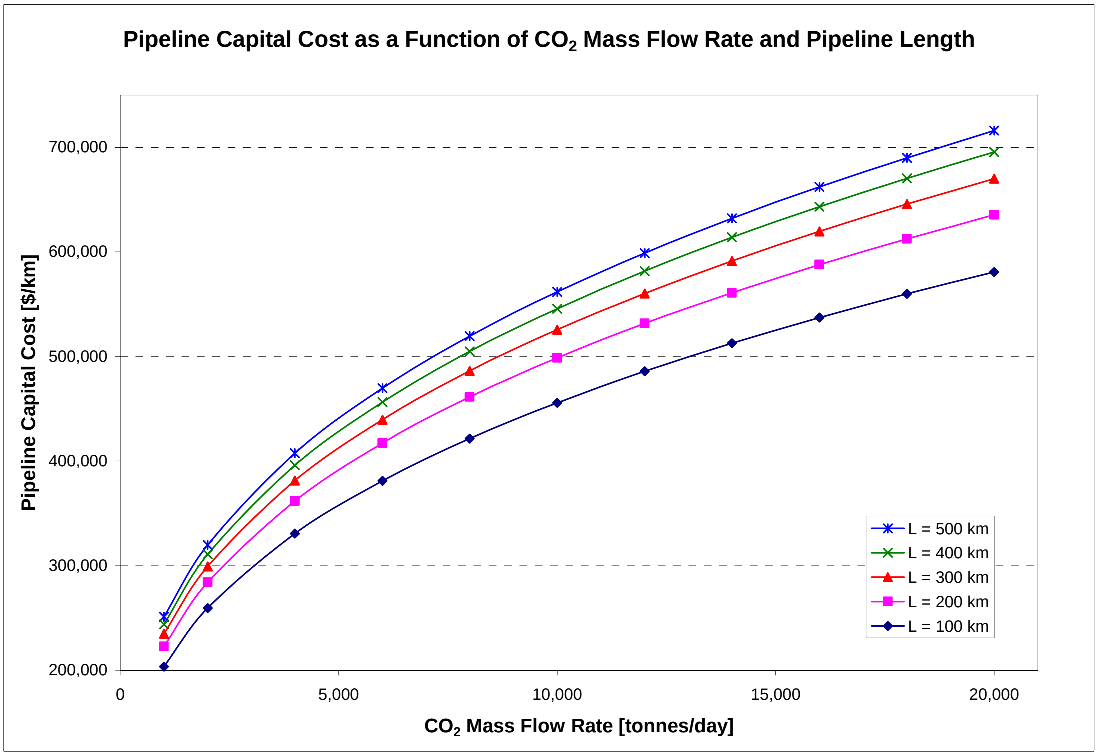


Figure 6: Pipeline Capital Cost as a Function of CO_2 Mass Flow Rate and Pipeline Length

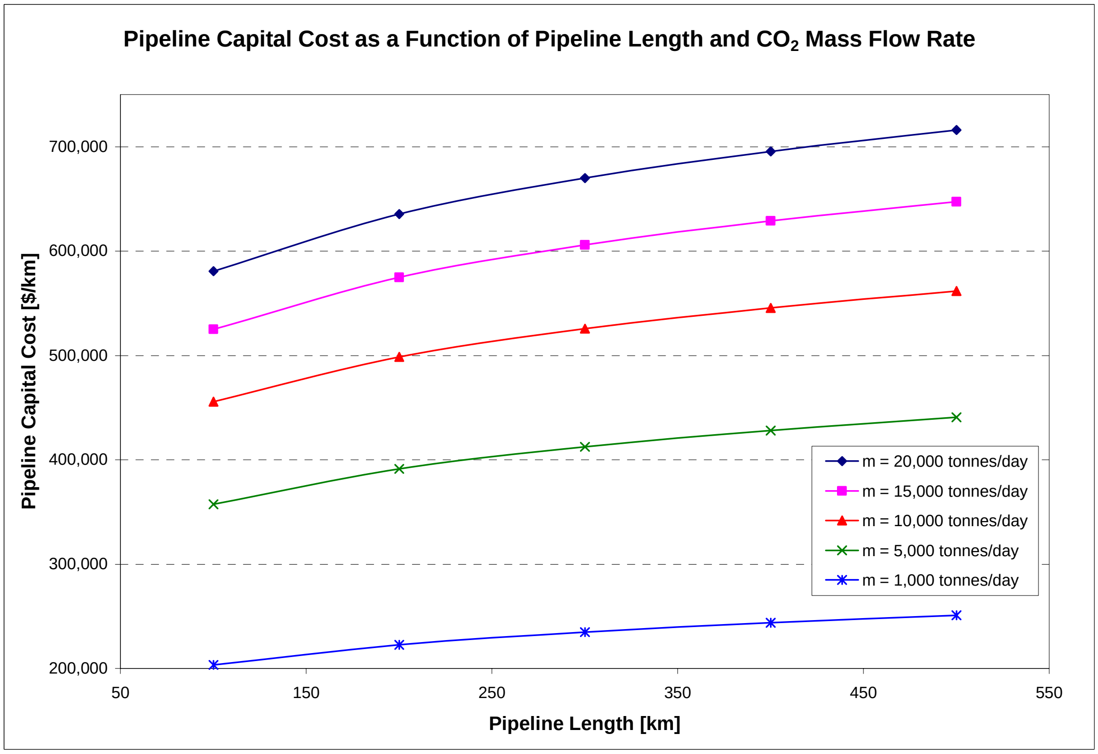


Figure 7: Pipeline Capital Cost as a Function of Pipeline Length and CO_2 Mass Flow Rate

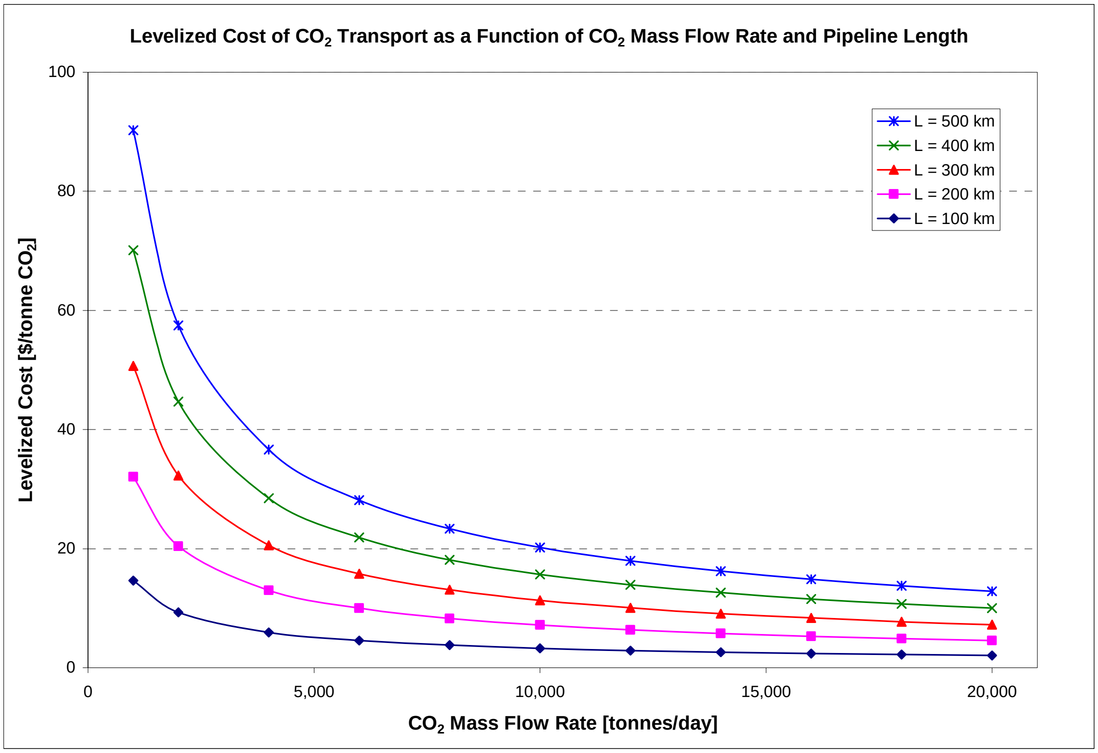


Figure 8: Levelized Cost of CO_2 Transport as a Function of CO_2 Mass Flow Rate and Pipeline Length (F_L = 1.0 assumed; and F_T = 1.20 assumed as an approximate average of all terrains)

PART III: CO_2 INJECTION & STORAGE

Nomenclature

m = CO_2 mass flow delivered to injection site per day [tonnes/day]
m_{year} = CO_2 mass flow delivered to injection site per year [tonnes/year]
CF = capacity factor [-]
P_{sur} = surface pressure of CO_2 at the top of the injection well [MPa]
P_{res} = pressure in the reservoir [MPa]
P_{down} = downhole injection pressure of CO_2 (i.e., pressure at bottom of injection well) [MPa]
P_{inter} = average between reservoir pressure (P_{res}) and downhole injection pressure (P_{down}) [MPa]
ΔP_{down} = downhole pressure difference = $P_{down} - P_{res}$ [MPa]
T_{sur} = surface temperature of CO_2 at the top of the injection well [^oC]
G_g = geothermal gradient [$^oC/km$]
T_{res} = temperature in the reservoir [^oC]
d = reservoir depth [m]
h = reservoir thickness [m]
k_a = absolute permeability of reservoir [millidarcy (md)]
k_v = vertical permeability of reservoir [millidarcy (md)]
k_h = horizontal permeability of reservoir [millidarcy (md)]
μ_{inter} = CO_2 viscosity at intermediate pressure (P_{inter}) [mPa-s]
μ_{sur} = CO_2 viscosity at surface temperature (T_{sur}) [Pa-s]
ρ_{sur} = CO_2 density at surface temperature (T_{sur}) and surface pressure (P_{sur}) [kg/m^3]
CO_2 mobility = absolute permeability (k_a) divided by CO_2 viscosity (μ_{inter}) [md/mPa-s]
CO_2 injectivity = mass flow rate of CO_2 that can be injected per unit of reservoir thickness (h) and per unit of downhole pressure difference ($P_{down} - P_{res}$) [tonnes/day/m/MPa]
g = gravitational constant [m/s^2]
P_{grav} = gravity head of CO_2 column in injection well [MPa]
D_{pipe} = injection pipe diameter [m]
Re = Reynold's number [-]
ε = injection pipe roughness factor [ft]
F_f = Fanning friction factor [-]
v_{pipe} = CO_2 velocity in injection pipe [m/s]
ΔP_{pipe} = frictional pressure loss in injection pipe [MPa]
$Q_{CO2/well}$ = CO_2 injection rate per well [tonnes/day/well]
N_{calc} = calculated number of injection wells [-]
N_{well} = actual number of injection wells (i.e., rounded up to nearest integer) [-]
C_{site} = capital cost of site screening and evaluation [$]
C_{equip} = capital cost of injection equipment [$]
C_{drill} = capital cost for drilling of the injection well [$]
C_{total} = total capital cost of injection wells [$]
C_{annual} = annualized capital cost of injection wells [$/yr]
CRF = Capital Recovery factor [-/yr]
$O\&M_{daily}$ = O&M costs due to normal daily expenses [$/yr]
$O\&M_{cons}$ = O&M costs due to consumables [$/yr]
$O\&M_{sur}$ = O&M costs due to surface maintenance [$/yr]

$O\&M_{subsur}$ = O&M costs due to subsurface maintenance [$/yr]
$O\&M_{total}$ = total O&M costs [$/yr]

Injection Well Number Calculation

The number of CO_2 injection wells that are required is strongly dependent on the properties of the particular geological reservoir that is being used to store the CO_2. Every reservoir is unique, however, and reservoir properties are quite varied. MIT [8] has done some statistical analysis on properties of actual reservoirs in the U.S., and they subsequently use the ranges in the following tables for their study on CO_2 storage in saline aquifers and in gas and oil reservoirs. The properties shown are reservoir pressure (P_{res}), thickness (h), depth (d), and horizontal permeability (k_h).

Parameter	Units	Aquifer Base Case	Aquifer High Cost Case	Aquifer Low Cost Case
Pressure	MPa	8.4	11.8	5.0
Thickness	m	171	42	703
Depth	m	1,239	1,784	694
Permeability	md	22	0.8	585

Table 1: Representative Range of Saline Aquifer Reservoir Properties [8]

Parameter	Units	Oil Reservoir Base Case	Oil Reservoir High Cost Case	Oil Reservoir Low Cost Case
Pressure	MPa	13.8	20.7	3.5
Thickness	m	43	21	61
Depth	m	1,554	2,134	1,524
Permeability	md	5	5	19

Table 2: Representative Range of Oil Reservoir Properties [8]

Parameter	Units	Gas Reservoir Base Case	Gas Reservoir High Cost Case	Gas Reservoir Low Cost Case
Pressure	MPa	3.5	6.9	2.1
Thickness	m	31	15	61
Depth	m	1,524	3,048	610
Permeability	md	1	0.8	10

Table 3: Representative Range of Gas Reservoir Properties [8]

The reservoir properties corresponding to "High Cost Case" in the preceding tables can be taken as the values that will lead to the maximum number of injection wells and, thus, maximum costs. Similarly, the "Low Cost Case" values will lead to the minimum costs. The "Base Case" values can be taken as statistically representative of any one reservoir.

By assuming a surface temperature of 15 ^{o}C (i.e., at the top of the injection well) and a geothermal gradient of 25 ^{o}C/km [8], and taking reservoir depth (d) from the above tables, the reservoir temperature can be approximated.

$T_{res} = T_{sur} + d*(G_g / 1000)$ (where $T_{sur} = 15\ ^oC$ and $G_g = 25\ ^oC/km$)

The procedure for calculating the number of CO_2 injection wells is iterative. To begin, one must assume a value for the downhole injection pressure (P_{down}), which is the CO_2 pressure at the bottom of the injection well. A reasonable first approximation for P_{down} is 17 MPa. The intermediate pressure of CO_2 in the reservoir (P_{inter}) is the average between the downhole injection pressure (P_{down}) and the reservoir pressure far from the injection well (P_{res}), which is taken from the above tables.

$$P_{inter} = (P_{down} + P_{res}) / 2$$

Based on P_{inter}, the CO_2 viscosity in the reservoir near the bottom of the injection well (μ_{inter}) can be approximated. As stated in the previous section, since CO_2 exhibits unusual trends in its properties over the range of temperatures and pressures that would be experienced with injection and storage, it is difficult to provide a single value for viscosity here. Therefore, the reader is referred to either of two CO_2 property websites, [5] and [6], or to the set of correlation equations of McCollum [7]. Each of these references provide an easy way of obtaining CO_2 density and viscosity if one knows only two basic parameters—temperature and pressure.

The absolute permeability of the reservoir (k_a) is found by an equation from [13].

$k_a = (k_h * k_v)^{0.5} = (k_h * 0.3k_h)^{0.5}$ (where k_h is taken from the above tables)

The mobility of CO_2 in the reservoir is thus [8]:

$$CO_2 \text{ mobility} = k_a / \mu_{inter}$$

The injectivity of CO_2 is then found by [13]:

$$CO_2 \text{ injectivity} = 0.0208 * CO_2 \text{ mobility}$$

And the CO_2 injection rate per well is calculated by the following equation [8].

$$Q_{CO2/well} = (CO_2 \text{ injectivity}) * h * \Delta P_{down}$$
$$= (CO_2 \text{ injectivity}) * h * (P_{down} - P_{res})$$ (where h is taken from the above tables)

The number of injection wells is based on the flow rate of CO_2 that is delivered to the injection site and the injection rate per well [8].

$$N_{calc} = m / Q_{CO2/well}$$

This is the calculated number of injection wells, not the actual number. The actual number of wells must, of course, be an integer value and will be determined in the final step.

As stated previously, the calculation of well number is iterative, due to the downhole injection pressure (P_{down}) initially being unknown. P_{down} is simply the pressure increase due to the gravity head of the CO_2 column in the injection well (P_{grav}), accounting for the fact that there is some pressure drop due to friction in the injection pipe (ΔP_{pipe}) [8].

$P_{down} = P_{sur} + P_{grav} - \Delta P_{pipe}$

The gravity head is a function of the gravitational constant (g) and the density of CO_2 (ρ_{sur}) at the surface temperature (T_{sur}) and surface pressure (P_{sur}). Once again, for estimating CO_2 density the reader is referred to either of two CO_2 property websites, [5] and [6], or to the set of correlation equations of McCollum [7].

$P_{grav} = (\rho_{sur} * g * d) / 10^6$ (where $g = 9.81\ m/s^2$)

The frictional pressure loss in the injection pipe is found in much the same way as the pipeline diameter was calculated in a previous section of this report. The Reynold's number (Re) is first found by the following equation, adapted from [8]:

$Re = 4 * (m*1000/24/3600/N_{calc}) / \pi / \mu_{sur} / D_{pipe}$

(where 1000, 24, and 3600 are unit conversion factors)

The CO_2 viscosity (μ_{sur}) at the surface temperature (T_{sur}) can be approximated by [5], [6], or [7]. The injection pipe diameter (D_{pipe}) is assumed to be one of the following values, based on MIT's report [8]:

- 0.059 m (~2.3 in) for all cases except the aquifer base case and aquifer low cost case;
- 0.1 m (~3.9 in) for the aquifer base case;
- 0.5 m (19.7 in) for the aquifer low cost case (Though, the MIT report mentions that an injection pipe of this size is too large to be used in practice. Therefore, a diameter of 0.12 m (~4.7 in) is assumed to be a reasonable upper limit.)

The Fanning friction factor (F_f) for flow in the injection pipe is calculated by the following equation from [14]:

$$F_f = \frac{1}{4\left[-1.8\log_{10}\left\{\frac{6.91}{\text{Re}} + \left(\frac{0.3048\,(\varepsilon / D_{pipe})}{3.7}\right)^{1.11}\right\}\right]^2}$$

(where $\varepsilon = 0.00015$ ft is assumed by [8])

The frictional pressure drop is then calculated based on the CO_2 velocity in the injection pipe (v_{pipe}) [14].

$v_{pipe} = (m*1000/24/3600/N_{calc}) / (\rho_{sur} * \pi * (D_{pipe}/2)^2)$

$\Delta P_{pipe} = (\rho_{sur}*g*F_f*d*v_{pipe}^2) / (D_{pipe}*2*g) / 10^6$

Once again, the downhole injection pressure (P_{down}) is calculated by:

$P_{down} = P_{sur} + P_{grav} - \Delta P_{pipe}$

This calculated value for P_{down} can now be used to begin another iteration. The iterative process for calculating P_{down} should be carried out over and over again until there is very little difference (i.e., < 1%) between iterations.

Once P_{down} is known, the actual number of injection wells (N_{well}) can be found by rounding the calculated number of wells (N_{calc})—from the final iteration—up to the nearest integer.

$N_{well} = \text{ROUND\_UP}\ (N_{calc})$

Capital, O&M, and Levelized Costs of CO_2 Injection & Storage

*** All costs are expressed in year 2005 US$

The capital cost of site screening and evaluation (C_{site}) has been scaled up into year 2005$ based on an estimate by Smith [15].

$C_{site} = 1{,}857{,}773$

Equations for estimating the capital cost of injection equipment were developed by the MIT report [8] based on actual injection well costs given by the Energy Information Administration (EIA) in their annual "Costs and Indices for Domestic Oil and Gas Field Equipment and Production Operations" report. Injection equipment costs include supply wells, plants, distribution lines, headers, and electrical services [16]. The equations of [8] have been scaled up into year 2005$.

$C_{equip} = N_{well} * \{49{,}433 * [m / (280*N_{well})]^{0.5}\}$

MIT also developed an equation for estimating the drilling cost of an onshore injection well based on data from the "1998 Joint American Survey (JAS) on Drilling Costs" report. The equations of [8] have been scaled up into year 2005$.

$C_{drill} = N_{well} * 10^6 * 0.1063e^{0.0008*d}$

Therefore, the total capital cost is given by:

$C_{total} = C_{site} + C_{equip} + C_{drill}$

The capital cost can be annualized by applying a capital recovery factor (CRF) of 0.15.

$C_{annual} = C_{total} * CRF$ (where CRF = 0.15/yr)

O&M costs were also developed from the EIA "Costs and Indices for Domestic Oil and Gas Field Equipment and Production Operations" report. They can be grouped into the following four categories: Normal Daily Expenses ($O\&M_{daily}$), Consumables ($O\&M_{cons}$), Surface Maintenance ($O\&M_{sur}$), and Subsurface Maintenance ($O\&M_{subsur}$). Costs have been scaled up into 2005$.

$O\&M_{daily} = N_{well} * 7{,}596$

$O\&M_{cons} = N_{well} * 20{,}295$

$O\&M_{sur} = N_{well} * \{15{,}420 * [m / (280*N_{well})]^{0.5}\}$

$O\&M_{subsur} = N_{well} * \{5669 * (d / 1219)\}$

$O\&M_{total} = O\&M_{daily} + O\&M_{cons} + O\&M_{sur} + O\&M_{subsur}$

The total annual costs are thus:

Total Annual Cost [$/yr] = $C_{annual} + O\&M_{total}$

The total amount of CO_2 that must be injected and stored every year is found by applying a capacity factor (CF) of 0.80.

$m_{year} = m * 365 * CF$ (where CF = 0.80)

Finally, the levelized cost of CO_2 injection and storage is given by:

Levelized Cost [$/tonne CO_2] = (Total Annual Cost) / m_{year}

The following graphs show the sensitivity of both the levelized costs and number of injection wells to a few of the parameters that could vary between CO_2 storage reservoirs. To be sure, carbon capture and sequestration is highly site specific, and the properties of different reservoirs may be wildly different. In the following graphs, for consistency we have used a common set of parameters, and depending on the particular graph, some parameters are held constant while one or two of the others are varied. The common parameters, for the most part, correspond to the Aquifer Base Case values highlighted above

A few things are worth mentioning with regard to the graphs. For starters, as one would expect, the levelized cost of CO_2 storage decreases as the amount of CO_2 to be sequestered increases—i.e., economies of scale are present. Conversely, more injection wells are required at higher flow rates. In addition, as the diameter of the injection pipe gets smaller, the number of injection wells must be increased to compensate, which translates into higher levelized costs at smaller diameters. Moreover, as the reservoir gets thicker and is more permeable, fewer injection wells are needed to do the same job. Note that reservoir depth and pressure were also examined in this sensitivity analysis, but it was found that the number of injection wells is not as dependent on these two parameters as it is for reservoir thickness and permeability. Thus, they have not been shown here.

Common Design Bases		
CO2 flow rate to injection field	1,000 to 20,000	tonnes/day
Plant Capacity Factor	0.80	
Surface pressure (pipeline outlet)	10.3	MPa
Surface temperature	15.0	C
Reservoir temperature	46.0	C
Reservoir depth	1239	m
Reservoir thickness	10 to 1000	m
Reservoir permeability (horizontal)	0.1 to 500	md
Reservoir pressure	8.4	MPa
Injection Pipe Diameter	0.059, 0.1, 0.15, or 0.2	m
Common Economic Bases		
Reference Year for Dollar	2005	
Project Lifetime	20	years
Discount Rate	0.10	

Table 4: Common Set of Parameters Used in Sensitivity Analysis

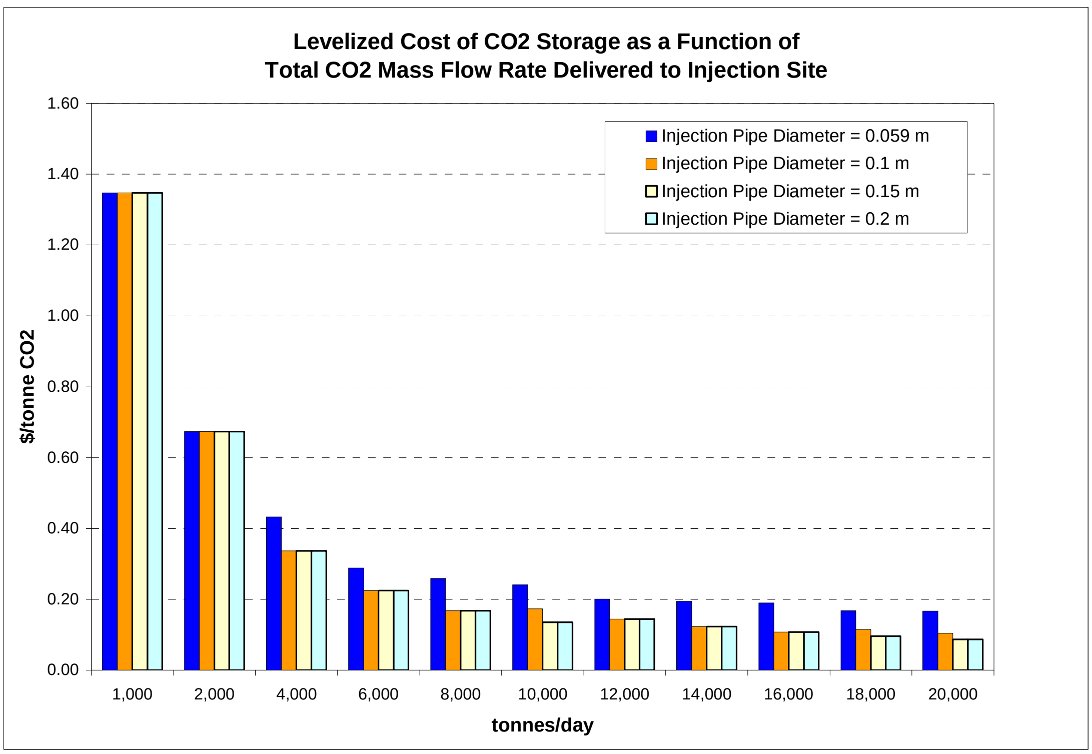


Figure 9: Levelized Cost of CO_2 Storage as a Function of Total CO_2 Mass Flow Rate Delivered to Injection Site

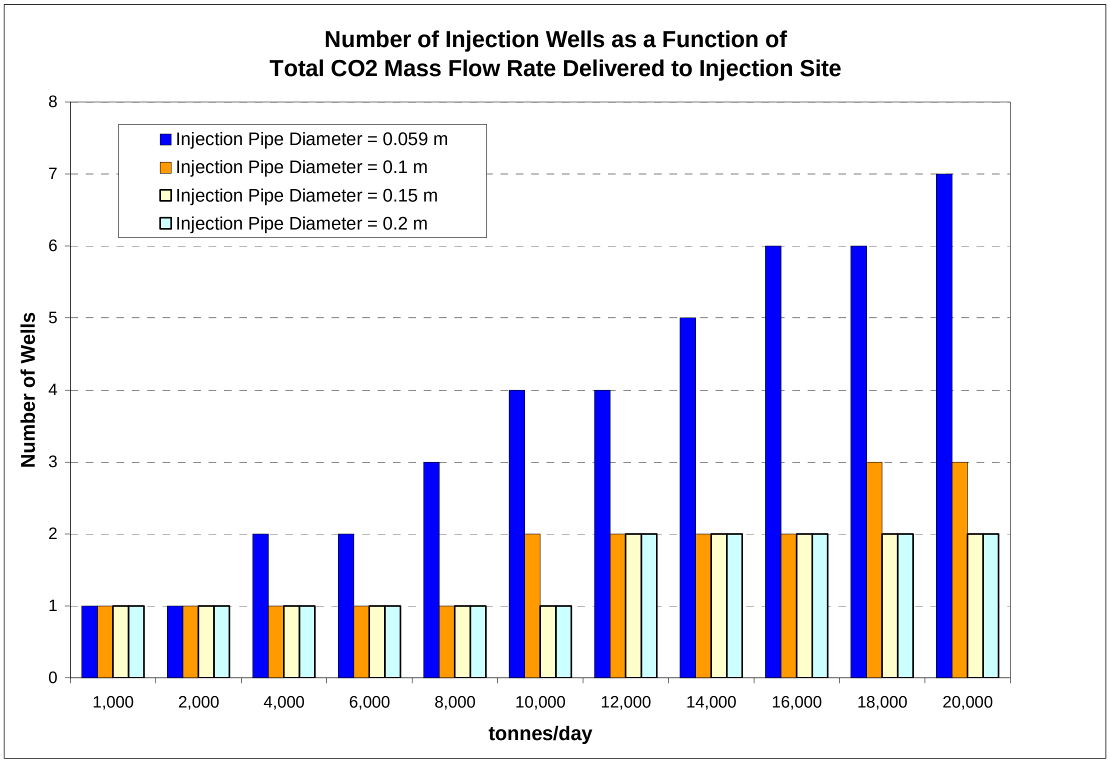


Figure 10: Number of Injection Wells as a Function of Total CO_2 Mass Flow Rate Delivered to Injection Site

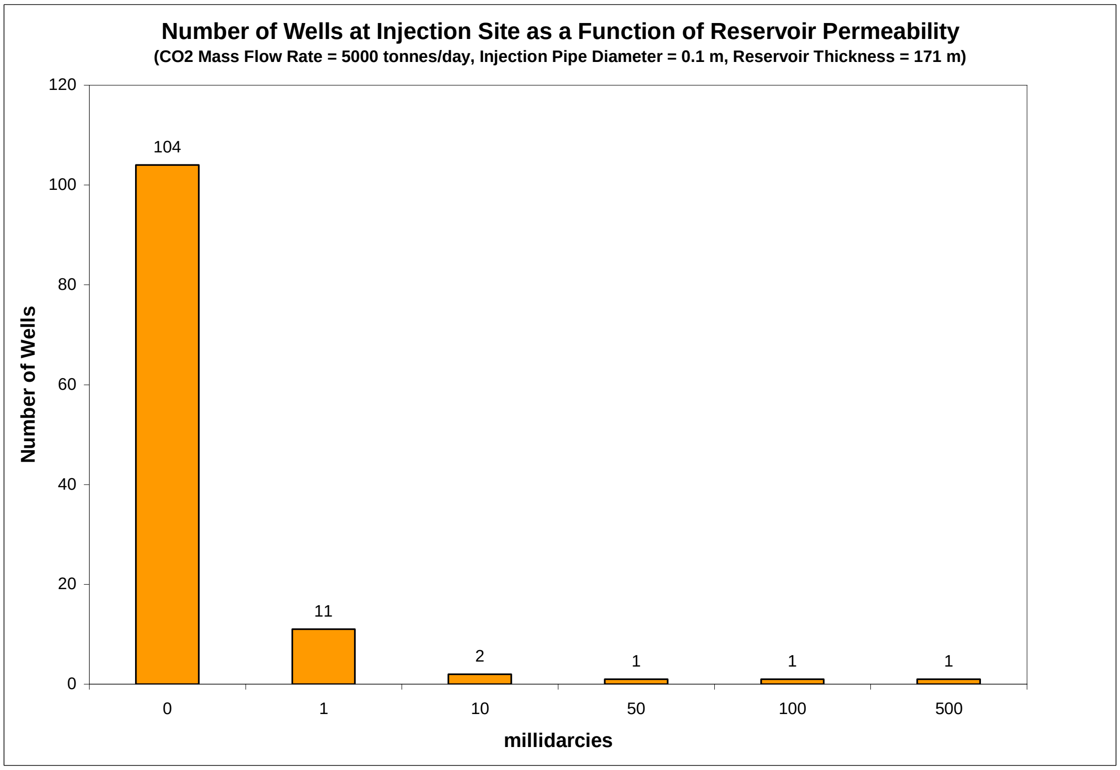


Figure 11: Number of Injection Wells as a Function of Reservoir Permeability

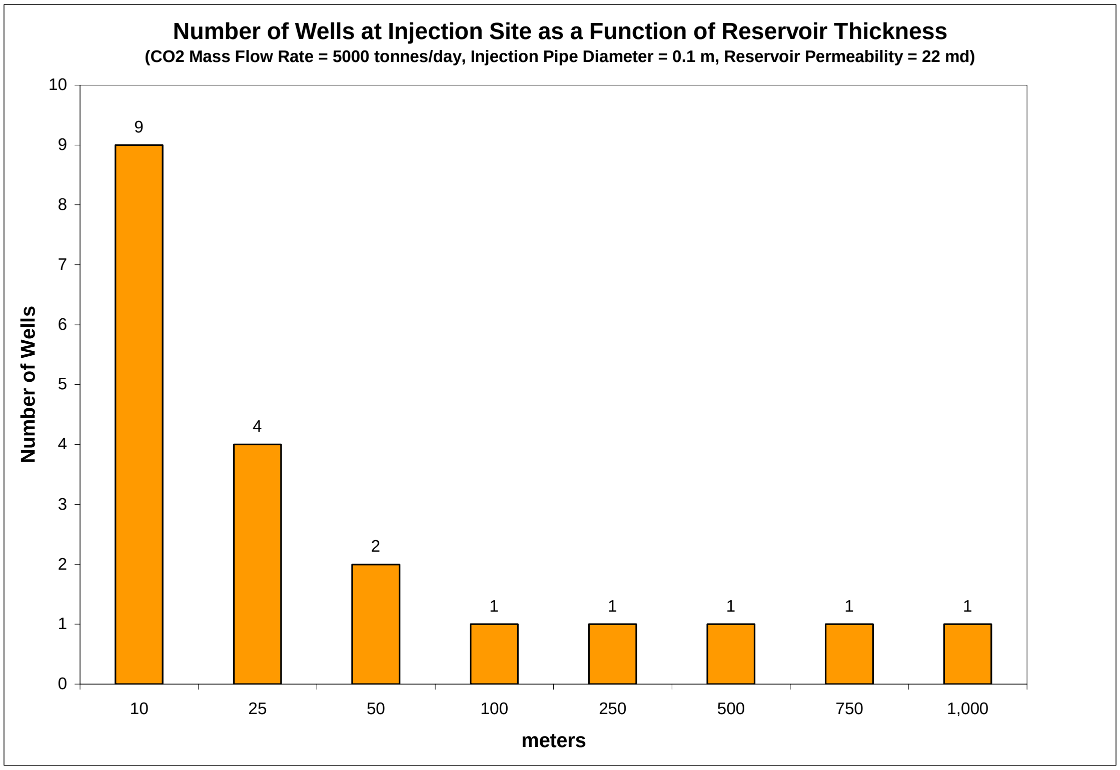


Figure 12: Number of Injection Wells as a Function of Reservoir Thickness

REFERENCES

[1] IEA Greenhouse Gas R&D Programme, "Transmission of CO2 and Energy," Report no. PH4/6 (March 2002).

[2] Mohitpour, M., H. Golshan, and A. Murray, "Pipeline Design & Construction: A Practical Approach, The American Society of Mechanical Engineers, New York (2000).

[3] "Practical Aspects of CO2 Flooding", Society of Petroleum Engineers (SPE) Monograph, Vol. 22, Appendix F, (2002).

[4] Hendriks, C., W. Graus, and F. van Bergen, "Global Carbon Dioxide Storage Potential and Costs", Ecofys report no. EEP-02001 (2004).

[5] NatCarb, US Department of Energy National Energy Technology Laboratory, http://www.natcarb.org/Calculators/co2_prop.html (Accessed on February 11, 2006).

[6] National Institute of Standards and Technology, http://webbook.nist.gov/chemistry/fluid/ (Accessed on February 11, 2006).

[7] McCollum, D.L., "Simple Correlations for Estimating Carbon Dioxide Density and Viscosity as a Function of Temperature and Pressure", Institute of Transportation Studies, University of California-Davis (2006).

[8] Herzog, Heddle, and Klett, "The Economics of CO2 Storage", MIT Laboratory for Energy and the Environment, Pub. # LFEE 2003-003 RP (August 2003).

[9] McCollum, D.L., "Comparing Techno-Economic Models for Pipeline Transport of Carbon Dioxide", Institute of Transportation Studies, University of California-Davis (2006).

[10] Hendriks, N., T. Wildenborg, P. Feron, W. Graus, R. Brandsma, "EC-Case Carbon Dioxide Sequestration," M70066, Ecofys (December 2003).

[11] IEA Greenhouse Gas R&D Programme, "Building the Cost Curves for CO_2 Storage: European Sector," Report no. 2005/2 (February 2005).

[12] IEA Greenhouse Gas R&D Programme, "Building the Cost Curves for CO_2 Storage: North America," Report no. 2005/3 (February 2005).

[13] Law, D. and S. Bachu, "Hydrogeological and numerical analysis of CO_2 disposal in deep aquifers in the Alberta sedimentary basin," Energy Conversion Mgmt., 37:6-8, pp. 1167-1174 (1996).

[14] Herzog, Howard, personal e-mail communication (February 2006).

[15] Smith, L.A. et al., "Engineering and Economic Assessment of Carbon Dioxide Sequestration in Saline Formations," presented at the First National Conference on Carbon Sequestration, Washington D.C. (May 14-17, 2001).

[16] "Costs and Indices for Domestic Oil and Gas Field Equipment and Production Operations" report, Energy Information Administration (June 2005).

[17] Kreutz, T., R. Williams, S. Consonni, and P. Chiesa, "Co-production of hydrogen, electricity and CO_2 from coal with commercially ready technology. Part B: Economic Analysis", International Journal of Hydrogen Energy, 30, pp. 769-784 (2005).

<u>SECTION II:</u>
Simple Correlations for Estimating Carbon Dioxide Density and Viscosity as a Function of Temperature and Pressure

David L. McCollum
dlmccollum@ucdavis.edu

Institute of Transportation Studies
University of California
One Shields Avenue
Davis, CA 95616

ABSTRACT

Recent years have seen an increased interest in carbon capture and sequestration (CCS)—the idea of capturing carbon dioxide (CO2) from the exhaust gases of power plants and industrial complexes, compressing the CO2 for pipeline transport, and finally injecting it underground in natural reservoirs, for example, saline aquifers and oil and gas wells. Engineers and researchers need to be able to estimate accurately the properties of CO2, a substance that exhibits unusual behavior in its properties. A number of equation of state correlations for estimating CO2's properties already exist, but these are often written in complex computer codes and are functions of a number of specific parameters that an inexperienced user might have trouble dealing with. This paper describes a set of simple correlations for estimating the density and viscosity of CO2 within the range of operating temperatures and pressures that might be encountered in CCS. The correlations are functions of only two input parameters: temperature and pressure. And since the correlation equations are based on experimentally-measured data, their agreement with reality, as well as with other correlations, is remarkable.

Keywords: carbon dioxide, CO2, CO_2, sequestration, pipeline, correlation, density, viscosity

DESCRIPTION OF CORRELATIONS

We have used experimentally-measured carbon dioxide (CO2) property data to develop a set of correlations for estimating the density and viscosity of CO2 over the range of operating temperatures and pressures that might be encountered in carbon capture and sequestration (CCS) applications. Specifically, we have limited our correlations to a temperature range of -1.1 to 82.2 ^{o}C (30 to 180 ^{o}F) and a pressure range of 7.6 to 24.8 MPa (1100 to 3600 psia), corresponding to the post-capture conditions of CO2 used in pipeline transport and underground injection. We obtain our experimental data from Kinder Morgan, a leading CO2 transporter in the United States [1]. We believe this data to be quite reliable, and apparently, so does the US Department of Energy National Energy Technology Laboratory's national carbon sequestration program, NatCarb, who also use the Kinder Morgan property data for their online CO2 property calculator [2]. The Kinder Morgan data gives a number of CO2's properties as functions of temperature and pressure; some examples include: density, viscosity, compressibility factor, heat capacity, enthalpy, entropy, phase, and so on. With this data, we simply plotted the density/viscosity vs. pressure for a given temperature and generated a sixth-order polynomial regression equation to best fit the data. We then repeated this procedure at all of the other temperature values that we had access to. Some example graphs are shown below.

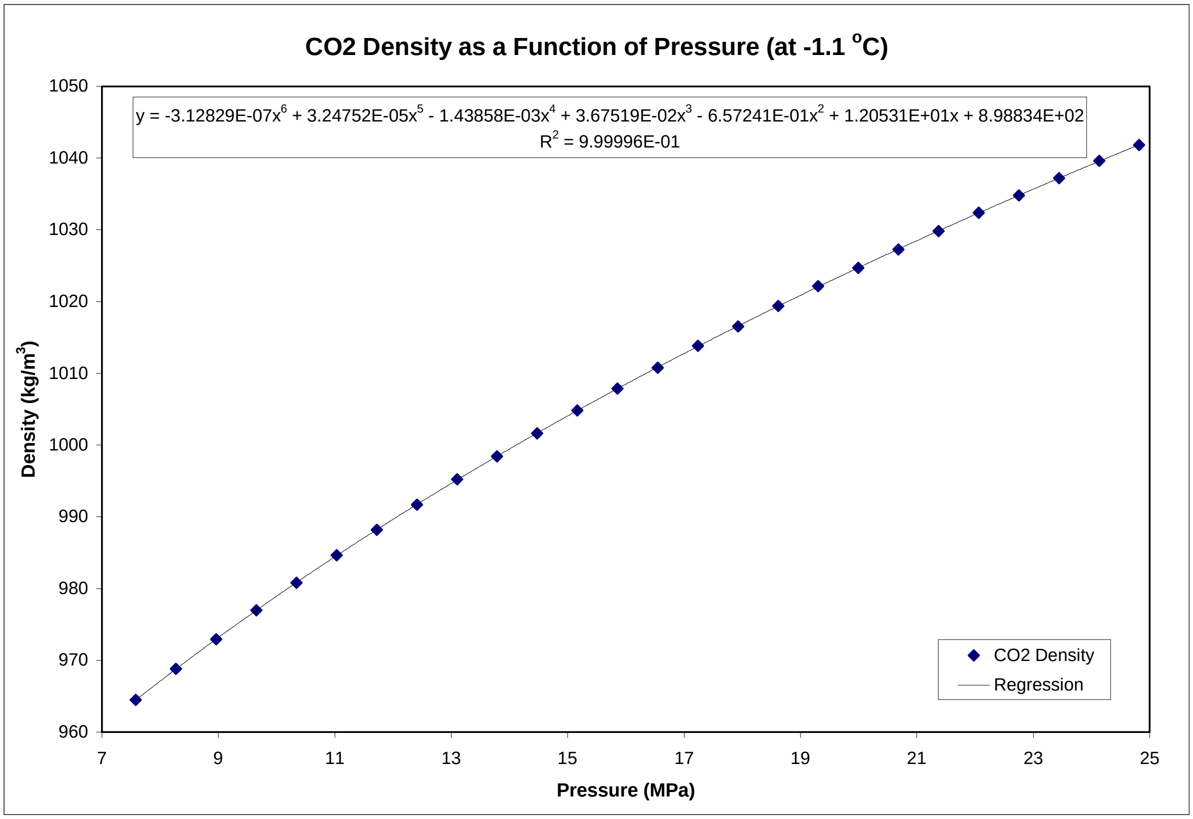


Figure 1: CO2 density as a function of pressure at -1.1 ^{o}C

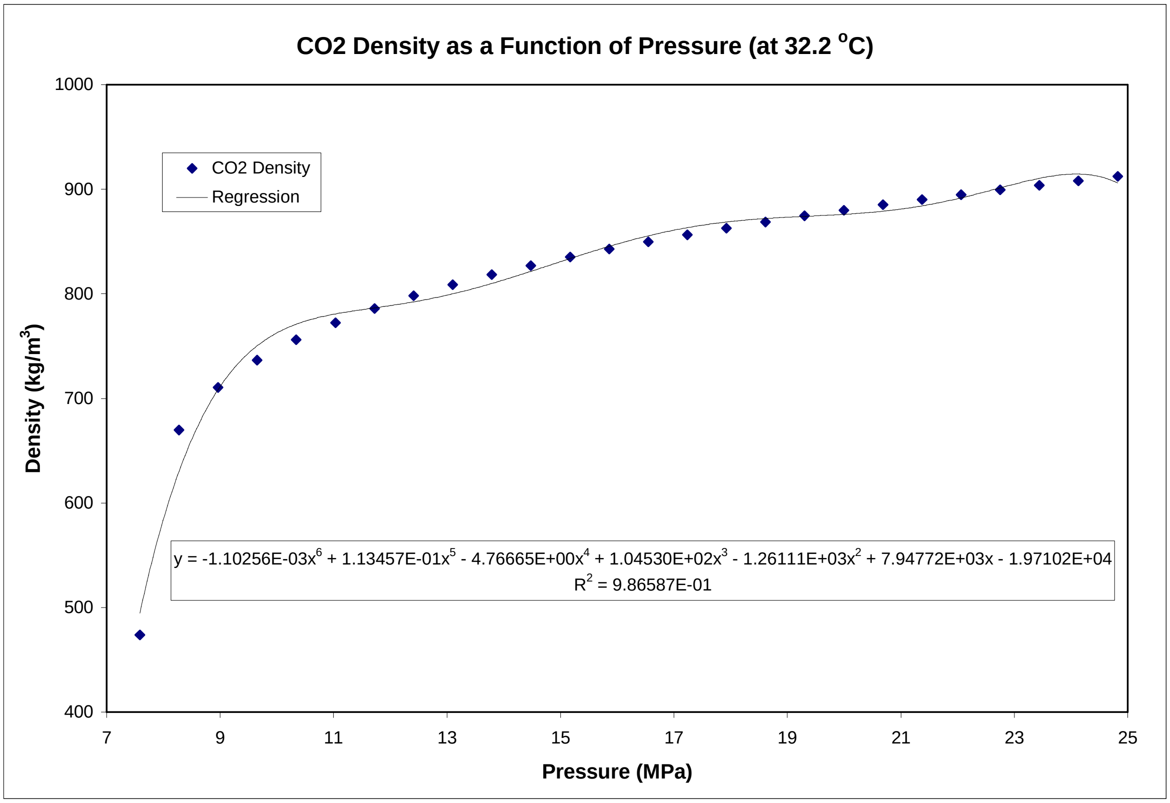


Figure 2: CO_2 density as a function of pressure at 32.2 °C

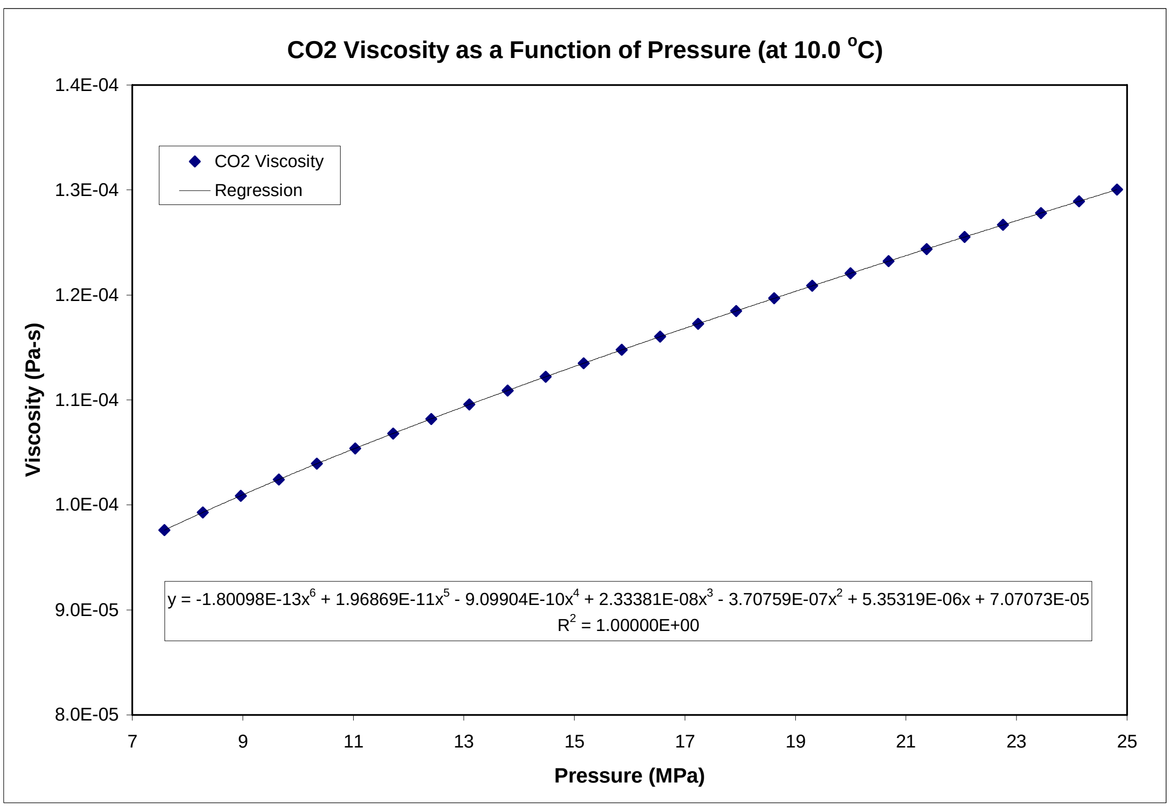


Figure 3: CO_2 viscosity as a function of pressure at 10.0 °C

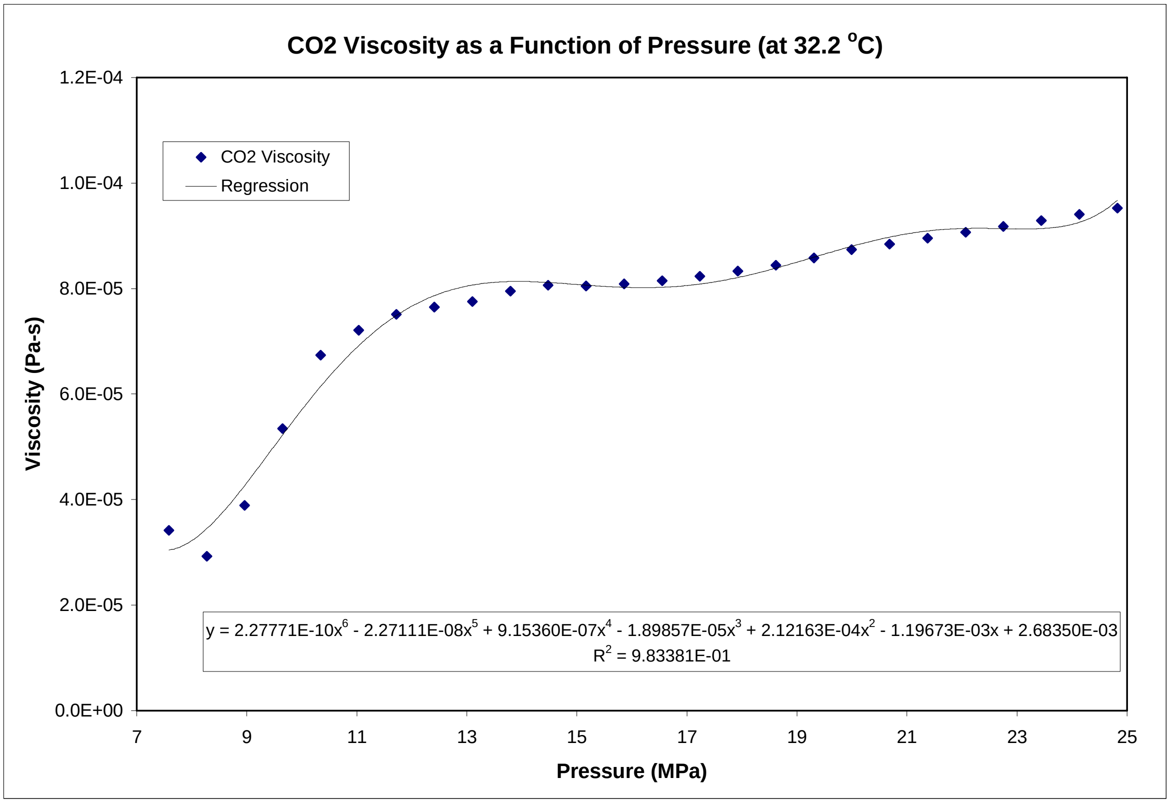


Figure 4: CO2 viscosity as a function of pressure at 32.2 °C

In total, we generated 32 graphs similar to the ones seen above (16 density graphs and 16 viscosity graphs for each of the 16 temperatures that we had access to). These particular four graphs are shown because they are representative of all of the others. On each of the graphs, the sixth-order polynomial regression equation and R^2 correlation coefficient are shown. The 'x' value in the regressions represents pressure (in MPa) and the 'y' value represents either density (in kg/m^3) or viscosity (in Pa-s). In general, the R^2 coefficient for all of the regressions, both density and viscosity, is greater than 0.995, showing excellent fit, except at temperatures just slightly above the critical temperature of CO2, 31.0 °C. (Note that all of pressure values considered here are above the critical pressure of CO2, 7.38 MPa.) But even at temperatures just slightly above the critical temperature, e.g. 32.2 °C, the R^2 coefficients for both density and viscosity are still greater than 0.983 (see Figures 2 and 4).

After generating all of the regression equations for density and viscosity at each of the given temperatures, we organized the regression equation coefficients into tabular form. In other words, for every temperature value there is a unique regression equation that relates pressure to either density or viscosity. Since each of these equations is unique, it has its own set of unique regression equation coefficients—i.e., the constants that precede the x^6, x^5, x^4, x^3, x^2, and x terms and the final constant term in the equations shown on the graphs above. These coefficients are shown for both density and viscosity in the tables below.

CO2 Density

Dependence of regression equation coefficients on temperature

Temperature (^oC)	Regression Equation Coefficient						
	a (x^6)	b (x^5)	c (x^4)	d (x^3)	e (x^2)	f (x)	g
-1.1	-3.12829E-07	3.24752E-05	-1.43858E-03	3.67519E-02	-6.57241E-01	1.20531E+01	8.98834E+02
4.4	-9.54845E-08	1.97920E-05	-1.41421E-03	5.06981E-02	-1.07669E+00	1.77109E+01	8.42753E+02
10.0	-6.99274E-07	8.56082E-05	-4.41249E-03	1.25510E-01	-2.19938E+00	2.81960E+01	7.68647E+02
15.6	-2.92964E-07	6.57269E-05	-4.75451E-03	1.67603E-01	-3.31969E+00	4.21135E+01	6.70554E+02
21.1	-7.86428E-06	8.72837E-04	-4.02787E-02	9.97669E-01	-1.42859E+01	1.21788E+02	3.84188E+02
26.7	-4.14913E-05	4.43672E-03	-1.95389E-01	4.55038E+00	-5.96084E+01	4.30173E+02	-5.36390E+02
32.2	-1.10256E-03	1.13457E-01	-4.76665E+00	1.04530E+02	-1.26111E+03	7.94772E+03	-1.97102E+04
37.8	-5.42882E-04	5.98138E-02	-2.70792E+00	6.44535E+01	-8.50922E+02	5.92597E+03	-1.63183E+04
43.3	9.60943E-04	-9.44447E-02	3.73493E+00	-7.54076E+01	8.07616E+02	-4.21227E+03	8.42194E+03
48.9	1.02964E-03	-1.05231E-01	4.36150E+00	-9.33059E+01	1.07660E+03	-6.23329E+03	1.42664E+04
54.4	4.91938E-04	-5.30672E-02	2.32907E+00	-5.29027E+01	6.48716E+02	-3.97202E+03	9.61309E+03
60.0	1.78281E-05	-5.25573E-03	3.79601E-01	-1.19952E+01	1.86161E+02	-1.32231E+03	3.60656E+03
65.6	-2.01381E-04	1.79337E-02	-6.14241E-01	9.95370E+00	-7.50237E+01	2.48324E+02	-1.20531E+02
71.1	-2.27250E-04	2.17674E-02	-8.25519E-01	1.56315E+01	-1.53782E+02	7.78805E+02	-1.49200E+03
76.7	-1.72335E-04	1.71075E-02	-6.76015E-01	1.34315E+01	-1.39949E+02	7.57756E+02	-1.56388E+03
82.2	-1.04002E-04	1.07058E-02	-4.38694E-01	9.02417E+00	-9.70390E+01	5.47454E+02	-1.15792E+03

Table 1: Regression equation coefficients for CO2 density

CO2 Viscosity

Dependence of regression equation coefficients on temperature

Temperature (^oC)	Regression Equation Coefficient						
	a (x^6)	b (x^5)	c (x^4)	d (x^3)	e (x^2)	f (x)	g
-1.1	-3.76516E-14	4.42744E-12	-2.21897E-10	6.35275E-09	-1.20061E-07	3.21247E-06	9.69913E-05
4.4	-4.13198E-14	5.05771E-12	-2.67210E-10	8.10161E-09	-1.59689E-07	3.68596E-06	8.53395E-05
10.0	-1.80098E-13	1.96869E-11	-9.09904E-10	2.33381E-08	-3.70759E-07	5.35319E-06	7.07073E-05
15.6	-3.83675E-13	4.25032E-11	-1.97443E-09	4.99914E-08	-7.54380E-07	8.42586E-06	5.17798E-05
21.1	-9.83505E-13	1.08507E-10	-4.97927E-09	1.22724E-07	-1.75059E-06	1.58647E-05	2.01512E-05
26.7	-4.04273E-12	4.32435E-10	-1.90732E-08	4.45698E-07	-5.87710E-06	4.39583E-05	-6.75597E-05
32.2	2.27771E-10	-2.27111E-08	9.15360E-07	-1.89857E-05	2.12163E-04	-1.19673E-03	2.68350E-03
37.8	9.44539E-11	-9.37386E-09	3.75251E-07	-7.70019E-06	8.44425E-05	-4.57587E-04	9.69405E-04
43.3	4.61459E-11	-4.64533E-09	1.89478E-07	-3.98321E-06	4.49854E-05	-2.50385E-04	5.50761E-04
48.9	2.17356E-11	-2.27268E-09	9.72054E-08	-2.16667E-06	2.62433E-05	-1.57279E-04	3.81014E-04
54.4	1.75118E-11	-1.83939E-09	7.90905E-08	-1.77644E-06	2.17839E-05	-1.32903E-04	3.32020E-04
60.0	1.59447E-11	-1.66290E-09	7.09018E-08	-1.57981E-06	1.92861E-05	-1.17925E-04	2.99069E-04
65.6	1.33132E-11	-1.38244E-09	5.86429E-08	-1.30108E-06	1.58745E-05	-9.74570E-05	2.52370E-04
71.1	9.59612E-12	-9.94594E-10	4.21212E-08	-9.35052E-07	1.14752E-05	-7.09785E-05	1.90487E-04
76.7	4.94000E-12	-5.14144E-10	2.19389E-08	-4.94382E-07	6.23334E-06	-3.93456E-05	1.15441E-04
82.2	8.35493E-13	-9.23510E-11	4.29135E-09	-1.10162E-07	1.66420E-06	-1.16755E-05	4.94127E-05

Table 2: Regression equation coefficients for CO2 viscosity

With the above regression equation coefficients, the density and viscosity of CO2 at any temperature and pressure in the above ranges (-1.1 to 82.2 ^oC and 7.6 to 24.8 MPa) can easily and reliably be calculated. One word of caution, however, is not to use the coefficients to try and extrapolate beyond the above ranges, as this will surely generate inaccurate output. The calculation is outlined below in a series of steps.

1) Specify the operating temperature, T_{op} (in ^oC).

2) In the above regression coefficient tables for both density and viscosity, find the range of temperatures (T_{high} and T_{low}) that the operating temperature (T_{op}) is between.

3) In the above regression coefficient tables for both density and viscosity, find the regression equation coefficients that correspond to T_{high} and T_{low}—a, b, c, d, e, f, and g.

4) Specify the operating pressure, P_{op} (in MPa).

5) With P_{op} calculate the density at T_{high} and at T_{low} and the viscosity at T_{high} and at T_{low}. The following generic equation can be used to calculate ρ_{high}, ρ_{low}, μ_{high}, and μ_{low}:

$$\rho \text{ or } \mu = a*P_{op}^6 + b*P_{op}^5 + c*P_{op}^4 + d*P_{op}^3 + e*P_{op}^2 + f*P_{op} + g$$

6) Interpolate for ρ_{op} and μ_{op} by the following equations.

$$\rho_{op} = \{(\rho_{high} - \rho_{low}) * (T_{op} - T_{low}) / (T_{high} - T_{low})\} + \rho_{low}$$

$$\mu_{op} = \{(\mu_{high} - \mu_{low}) * (T_{op} - T_{low}) / (T_{high} - T_{low})\} + \mu_{low}$$

*** A simple example should serve to illustrate this calculation procedure.

1) Assume $T_{op} = 47.0$ ^oC

2) From the regression coefficient tables, we find that $T_{op} = 47.0$ ^oC is between $T_{high} = 48.9$ ^oC and $T_{low} = 43.3$ ^oC.

3) From the density and viscosity tables, the regression equation coefficients are:

Density (ρ):

T_{high}: a = 1.02964E-03, b = -1.05231E-01, c = 4.36150E+00, d = -9.33059E+01, e = 1.07660E+03, f = -6.23329E+03, g = 1.42664E+04

T_{low}: a = 9.60943E-04, b = -9.44447E-02, c = 3.73493E+00, d = -7.54076E+01, e = 8.07616E+02, f = -4.21227E+03, g = 8.42194E+03

Viscosity (μ):

T_{high}: a = 2.17356E-11, b = -2.27268E-09, c = 9.72054E-08, d = -2.16667E-06, e = 2.62433E-05, f = -1.57279E-04, g = 3.81014E-04

T_{low}: a = 4.61459E-11, b = -4.64533E-09, c = 1.89478E-07, d = -3.98321E-06, e = 4.49854E-05, f = -2.50385E-04, g = 5.50761E-04

4) Assume P_{op} = 10 MPa.

5) Calculate ρ_{high}, ρ_{low}, μ_{high}, and μ_{low}:

$$\rho_{high} = 1.02964E\text{-}03*P_{op}^6 + -1.05231E\text{-}01*P_{op}^5 + 4.36150E{+}00*P_{op}^4 + -9.33059E{+}01*P_{op}^3 + 1.07660E{+}03*P_{op}^2 + -6.23329E{+}03*P_{op} + 1.42664E{+}04 = 409.1 \text{ kg/m}^3$$

$$\rho_{low} = 9.60943E\text{-}04*P_{op}^6 + -9.44447E\text{-}02*P_{op}^5 + 3.73493E{+}00*P_{op}^4 + -7.54076E{+}01*P_{op}^3 + 8.07616E{+}02*P_{op}^2 + -4.21227E{+}03*P_{op} + 8.42194E{+}03 = 519.0 \text{ kg/m}^3$$

$$\mu_{high} = 2.17356E\text{-}11*P_{op}^6 + -2.27268E\text{-}09*P_{op}^5 + 9.72054E\text{-}08*P_{op}^4 + -2.16667E\text{-}06*P_{op}^3 + 2.62433E\text{-}05*P_{op}^2 + -1.57279E\text{-}04*P_{op} + 3.81014E\text{-}04 = 3.24E\text{-}05 \text{ Pa-s}$$

$$\mu_{low} = 4.61459E\text{-}11*P_{op}^6 + -4.64533E\text{-}09*P_{op}^5 + 1.89478E\text{-}07*P_{op}^4 + -3.98321E\text{-}06*P_{op}^3 + 4.49854E\text{-}05*P_{op}^2 + -2.50385E\text{-}04*P_{op} + 5.50761E\text{-}04 = 3.86E\text{-}05 \text{ Pa-s}$$

6) Interpolate for ρ_{op} and μ_{op}.

$$\rho_{op} = \{(409.1 - 519.0) * (47.0 - 43.3) / (48.9 - 43.3)\} + 519.0 = 446.4 \text{ kg/m}^3$$

$$\mu_{op} = \{(3.24E\text{-}05 - 3.86E\text{-}05) * (47.0 - 43.3) / (48.9 - 43.3)\} + 3.86E\text{-}05 = 3.45E\text{-}05 \text{ Pa-s}$$

COMPARISON WITH OTHER CORRELATIONS

We have compared our CO_2 correlations to other, more complex equation of state correlations and find that ours are in agreement. Garcia [3] does a nice job of explaining equation of state CO_2 property correlations and then comparing densities calculated by various correlations over a small range of temperatures and pressures. He provides the following comparison table:

P [bar]	A	B	C	D	E	F
80	233.86	231.33	232.21	232.21	223.64	234.10
100	447.03	446.78	449.77	449.77	465.68	505.36
120	632.51	632.03	633.86	633.86	622.76	624.86
160	747.38	746.09	747.32	747.32	734.66	725.30
200	803.21	802.24	803.11	803.24	798.04	784.54
300	883.39	883.01	883.73	883.78	895.91	877.51

A: Table values from Vargaftik et al. [1996]

B: Calculated values from Span and Wagner [1996]

C: Table values from Angus et al. [1976]

D: Calculated values from Altunin [1975]

E: Redlich Kwong with Morris and Turek [1986] parameters

F: Redlich Kwong with Spycher et al. [2003] parameters

Table 3: CO_2 density (kg/m^3) at 320 K (47.0 ^{o}C) as a function of temperature and pressure by various correlations

Note that all of the densities in the above table are for 320 K (i.e., ~47.0 ^{o}C). Now, look at the row of CO_2 densities that correspond to 100 bar (i.e., 10 MPa). By design, these are exactly the operating conditions that we used in the example above to illustrate our methods and equations. At these conditions, our correlations estimate the CO_2 density to be 446.4 kg/m^3, which is well within the range of values (446.78 – 505.36 kg/m^3) calculated by other, more complex correlations, as shown in Garcia's table. In addition, our correlations match well with online CO_2 property calculators like those of NatCarb and the National Institute of Standards and Technology [2, 4]. As previously mentioned, the NatCarb calculator uses the same Kinder Morgan property data that we use. The NIST calculator on the other hand uses the correlations of Span and Wagner [5], which Garcia references in his table above.

Density and viscosity values (both experimentally-measured and those calculated by our regression equations) are shown in the appendices for all of the temperature and pressure operating points that we had data for. Also in the appendix, we show the percent differences between the calculated and experimentally-measured values for both density and viscosity. In almost all cases, the percent difference is less than 1%, and much of the time it is less than 0.1%. The greatest differences occur near the critical point of CO2, with differences as high as 13.8% for density and 18.1% for viscosity. Therefore, if one is interested in designing a system where the temperature and pressure are near the critical point of CO2 (31.0 ^{o}C and 7.38 MPa) for much of the time, then perhaps a more complex equation of state CO2 property correlation should be used. But at virtually any other operating conditions (at least in the range of conditions studied here), our correlations provide very reliable results.

CONCLUSION

We have used experimentally-measured CO2 property data to create sixth-order polynomial correlation equations for estimating the density and viscosity of CO2. Our correlations are functions of only two parameters—temperature and pressure—and can be used in the range of -1.1 to 82.2 ^oC (30 to 180 ^oF) and 7.6 to 24.8 MPa (1100 to 3600 psia). In the case of carbon capture and sequestration, these operating ranges correspond to the post-capture/post-compression conditions of CO2 used in pipeline transport and underground injection. Our correlations provide a simple alternative to the more complex equation of state correlations that are often used. While these more complex correlations may provide slightly more accurate density and viscosity estimates near the critical point of CO2, for the vast majority of operating temperatures and pressures in the ranges mentioned above, our correlations are just as accurate and reliable and should be used with confidence. We believe that simple correlations of this kind will be demanded more and more in the future, as CCS continues to gain interest, especially among those engineers and researchers with little background in the field and who would prefer to use simple correlations to obtain accurate results.

** Note: Any parties interested in obtaining a copy of the Microsoft Excel file of the CO2 property correlations described in this report, should feel free to contact the author at dlmccollum@ucdavis.edu.*

REFERENCES

[1] "Practical Aspects of CO_2 Flooding", Society of Petroleum Engineers (SPE) Monograph, Vol. 22, Appendix F, (2002).

[2] NatCarb, US Department of Energy National Energy Technology Laboratory, http://www.natcarb.org/Calculators/co2_prop.html, Accessed on February 11, 2006.

[3] García, J.E., "Fluid Dynamics of Carbon Dioxide Disposal Into Saline Aquifers", PhD dissertation, University of California at Berkeley, Berkeley, California (December 2003).

[4] National Institute of Standards and Technology (NIST), http://webbook.nist.gov/chemistry/fluid/, Accessed on February 11, 2006.

[5] Span, R. and W. Wagner, "A New Equation of State for Carbon Dioxide Covering the Fluid Region from the Triple-Point Temperature to 1100 K at Pressures up to 800 MPa", J. Phys. Chem. Ref. Data, 25, 6, 1509-1596 (1996).

APPENDIX

Appendix 1: Regression equation coefficients for CO_2 density

Appendix 2: Regression equation coefficients for CO_2 viscosity

Appendix 3: CO_2 density as a function of temperature and pressure (experimentally-measured values from the Kinder Morgan property data)

Appendix 4: CO_2 density as a function of temperature and pressure (calculated values from the regression equations)

Appendix 5: Percent difference between the calculated and experimentally-measured density values at each of the temperature and pressure operating points

Appendix 6: CO_2 viscosity as a function of temperature and pressure (experimentally-measured values from the Kinder Morgan property data)

Appendix 7: CO_2 viscosity as a function of temperature and pressure (calculated values from the regression equations)

Appendix 8: Percent difference between the calculated and experimentally-measured viscosity values at each of the temperature and pressure operating points

CO2 Density

Dependence of regression equation coefficients on temperature

Temperature (^oC)	Regression Equation Coefficient						
	a (x^6)	b (x^5)	c (x^4)	d (x^3)	e (x^2)	f (x)	g
-1.1	-3.12829E-07	3.24752E-05	-1.43858E-03	3.67519E-02	-6.57241E-01	1.20531E+01	8.98834E+02
4.4	-9.54845E-08	1.97920E-05	-1.41421E-03	5.06981E-02	-1.07669E+00	1.77109E+01	8.42753E+02
10.0	-6.99274E-07	8.56082E-05	-4.41249E-03	1.25510E-01	-2.19938E+00	2.81960E+01	7.68647E+02
15.6	-2.92964E-07	6.57269E-05	-4.75451E-03	1.67603E-01	-3.31969E+00	4.21135E+01	6.70554E+02
21.1	-7.86428E-06	8.72837E-04	-4.02787E-02	9.97669E-01	-1.42859E+01	1.21788E+02	3.84188E+02
26.7	-4.14913E-05	4.43672E-03	-1.95389E-01	4.55038E+00	-5.96084E+01	4.30173E+02	-5.36390E+02
32.2	-1.10256E-03	1.13457E-01	-4.76665E+00	1.04530E+02	-1.26111E+03	7.94772E+03	-1.97102E+04
37.8	-5.42882E-04	5.98138E-02	-2.70792E+00	6.44535E+01	-8.50922E+02	5.92597E+03	-1.63183E+04
43.3	9.60943E-04	-9.44447E-02	3.73493E+00	-7.54076E+01	8.07616E+02	-4.21227E+03	8.42194E+03
48.9	1.02964E-03	-1.05231E-01	4.36150E+00	-9.33059E+01	1.07660E+03	-6.23329E+03	1.42664E+04
54.4	4.91938E-04	-5.30672E-02	2.32907E+00	-5.29027E+01	6.48716E+02	-3.97202E+03	9.61309E+03
60.0	1.78281E-05	-5.25573E-03	3.79601E-01	-1.19952E+01	1.86161E+02	-1.32231E+03	3.60656E+03
65.6	-2.01381E-04	1.79337E-02	-6.14241E-01	9.95370E+00	-7.50237E+01	2.48324E+02	-1.20531E+02
71.1	-2.27250E-04	2.17674E-02	-8.25519E-01	1.56315E+01	-1.53782E+02	7.78805E+02	-1.49200E+03
76.7	-1.72335E-04	1.71075E-02	-6.76015E-01	1.34315E+01	-1.39949E+02	7.57756E+02	-1.56388E+03
82.2	-1.04002E-04	1.07058E-02	-4.38694E-01	9.02417E+00	-9.70390E+01	5.47454E+02	-1.15792E+03

Appendix 1

CO2 Viscosity

Dependence of regression equation coefficients on temperature

Temperature (°C)	Regression Equation Coefficient						
	a (x^6)	b (x^5)	c (x^4)	d (x^3)	e (x^2)	f (x)	g
-1.1	-3.76516E-14	4.42744E-12	-2.21897E-10	6.35275E-09	-1.20061E-07	3.21247E-06	9.69913E-05
4.4	-4.13198E-14	5.05771E-12	-2.67210E-10	8.10161E-09	-1.59689E-07	3.68596E-06	8.53395E-05
10.0	-1.80098E-13	1.96869E-11	-9.09904E-10	2.33381E-08	-3.70759E-07	5.35319E-06	7.07073E-05
15.6	-3.83675E-13	4.25032E-11	-1.97443E-09	4.99914E-08	-7.54380E-07	8.42586E-06	5.17798E-05
21.1	-9.83505E-13	1.08507E-10	-4.97927E-09	1.22724E-07	-1.75059E-06	1.58647E-05	2.01512E-05
26.7	-4.04273E-12	4.32435E-10	-1.90732E-08	4.45698E-07	-5.87710E-06	4.39583E-05	-6.75597E-05
32.2	2.27771E-10	-2.27111E-08	9.15360E-07	-1.89857E-05	2.12163E-04	-1.19673E-03	2.68350E-03
37.8	9.44539E-11	-9.37386E-09	3.75251E-07	-7.70019E-06	8.44425E-05	-4.57587E-04	9.69405E-04
43.3	4.61459E-11	-4.64533E-09	1.89478E-07	-3.98321E-06	4.49854E-05	-2.50385E-04	5.50761E-04
48.9	2.17356E-11	-2.27268E-09	9.72054E-08	-2.16667E-06	2.62433E-05	-1.57279E-04	3.81014E-04
54.4	1.75118E-11	-1.83939E-09	7.90905E-08	-1.77644E-06	2.17839E-05	-1.32903E-04	3.32020E-04
60.0	1.59447E-11	-1.66290E-09	7.09018E-08	-1.57981E-06	1.92861E-05	-1.17925E-04	2.99069E-04
65.6	1.33132E-11	-1.38244E-09	5.86429E-08	-1.30108E-06	1.58745E-05	-9.74570E-05	2.52370E-04
71.1	9.59612E-12	-9.94594E-10	4.21212E-08	-9.35052E-07	1.14752E-05	-7.09785E-05	1.90487E-04
76.7	4.94000E-12	-5.14144E-10	2.19389E-08	-4.94382E-07	6.23334E-06	-3.93456E-05	1.15441E-04
82.2	8.35493E-13	-9.23510E-11	4.29135E-09	-1.10162E-07	1.66420E-06	-1.16755E-05	4.94127E-05

Appendix 2

CO_2 Density (kg/m^3) as a Function of Temperature (^oC) and Pressure (MPa)
(actual values from Kinder Morgan)

	Temperature (^oC)															
Pressure (MPa)	-1.1	4.4	10.0	15.6	21.1	26.7	32.2	37.8	43.3	48.9	54.4	60.0	65.6	71.1	76.7	82.2
7.6	964.5	933.1	898.2	858.0	808.5	739.1	473.8	254.4	220.9	200.9	186.8	175.7	166.8	159.2	152.7	146.9
8.3	968.8	938.4	904.9	866.9	821.6	763.3	669.6	371.3	274.7	239.6	218.0	202.6	190.5	180.7	172.4	165.3
9.0	973.0	943.5	911.1	874.9	833.0	781.4	710.4	577.0	361.7	290.3	255.7	233.4	217.1	204.2	193.7	184.9
9.7	977.0	948.3	917.1	882.5	843.1	796.3	736.5	648.4	489.7	359.8	302.1	269.1	246.8	230.0	216.7	205.8
10.3	980.8	952.9	922.7	889.5	852.2	808.9	756.2	686.7	582.9	447.9	359.5	311.1	280.5	258.5	241.7	228.1
11.0	984.7	957.3	928.0	895.9	860.4	820.0	772.4	713.1	634.2	528.0	425.6	359.6	318.3	289.9	268.8	252.1
11.7	988.2	961.4	932.9	902.0	868.0	829.9	786.0	733.6	668.1	584.5	490.0	412.5	359.9	324.1	297.8	277.6
12.4	991.7	965.6	937.7	907.8	875.1	838.9	798.0	750.5	693.6	624.1	543.2	465.0	404.1	360.6	328.7	304.5
13.1	995.2	969.4	942.2	913.2	881.7	847.1	808.6	764.7	713.8	653.7	584.4	511.8	447.9	398.4	361.1	332.7
13.8	998.4	973.3	946.7	918.3	887.9	854.7	818.2	777.4	730.8	677.4	616.6	551.2	488.6	435.7	394.2	361.7
14.5	1001.6	977.0	950.9	923.3	893.7	861.8	827.0	788.6	745.5	697.0	642.5	583.6	524.6	471.3	426.7	391.0
15.2	1004.8	980.5	955.0	928.0	899.3	868.5	835.2	798.7	758.3	713.6	664.1	610.6	555.8	503.8	458.1	420.0
15.9	1007.9	983.9	958.9	932.6	904.6	874.8	842.7	808.0	769.8	728.2	682.5	633.5	582.8	532.9	487.4	447.9
16.5	1010.8	987.2	962.7	936.9	909.7	880.7	849.8	816.5	780.4	741.2	698.7	653.2	606.0	558.9	514.4	474.3
17.2	1013.8	990.6	966.4	941.1	914.5	886.5	856.5	824.5	790.0	752.9	713.0	670.5	626.3	581.8	538.7	499.1
17.9	1016.5	993.8	969.9	945.1	919.1	891.7	862.8	831.8	798.8	763.6	725.8	685.9	644.3	602.1	560.8	521.9
18.6	1019.4	996.8	973.4	949.1	923.6	896.9	868.7	838.9	807.2	773.4	737.5	699.7	660.4	620.4	580.8	543.0
19.3	1022.1	999.9	976.8	952.9	928.0	901.8	874.4	845.5	814.9	782.3	748.2	712.2	674.9	636.9	598.9	562.2
20.0	1024.7	1002.8	980.0	956.6	932.1	906.6	879.7	851.7	822.1	790.8	758.0	723.7	688.2	651.8	615.4	580.0
20.7	1027.3	1005.6	983.2	960.1	936.1	911.1	885.0	857.6	829.0	798.8	767.3	734.3	700.3	665.6	630.6	596.4
21.4	1029.8	1008.5	986.4	963.7	940.0	915.5	890.0	863.2	835.4	806.2	775.8	744.1	711.4	678.1	644.6	611.4
22.1	1032.4	1011.2	989.5	967.0	943.8	919.8	894.8	868.7	841.6	813.3	783.8	753.4	721.8	689.8	657.4	625.4
22.8	1034.8	1014.0	992.3	970.2	947.5	923.8	899.4	874.0	847.5	820.0	791.5	761.8	731.6	700.5	669.4	638.3
23.4	1037.2	1016.5	995.4	973.4	951.0	927.8	903.8	878.9	853.1	826.4	798.5	770.0	740.5	710.7	680.5	650.4
24.1	1039.6	1019.1	998.1	976.6	954.5	931.6	908.1	883.7	858.4	832.3	805.4	777.7	749.2	720.2	690.9	661.7
24.8	1041.8	1021.7	1001.0	979.7	957.9	935.5	912.3	888.4	863.7	838.2	811.8	784.9	757.2	729.2	700.6	672.3

Appendix 3

CO_2 Density (kg/m^3) as a Function of Temperature (^oC) and Pressure (MPa)
(calculated values from regression equations)

	Temperature (^oC)													
Pressure (MPa)	-1.1	4.4	10.0	15.6	21.1	26.7	32.2	37.8	43.3	48.9	54.4	60.0	65.6	71.1
7.6	964.5	933.1	898.2	858.0	808.5	739.5	494.7	236.0	203.5	199.6	192.1	180.5	169.1	159.9
8.3	968.8	938.4	904.9	866.8	821.5	762.6	629.9	422.4	295.4	234.5	208.9	196.7	188.2	180.5
9.0	973.0	943.5	911.2	875.0	832.9	781.2	708.3	548.7	390.0	296.8	251.0	228.0	214.1	203.3
9.7	977.0	948.3	917.1	882.5	843.1	796.5	750.3	632.0	478.5	370.9	307.0	269.4	245.8	229.3
10.3	980.8	952.9	922.6	889.5	852.2	809.3	771.0	685.5	555.6	445.9	368.5	316.5	282.2	258.6
11.0	984.6	957.3	927.9	895.9	860.4	820.3	780.7	719.3	618.6	514.9	429.6	366.2	321.8	291.0
11.7	988.2	961.5	932.9	902.0	868.0	830.0	786.5	740.8	667.3	573.8	486.5	415.7	363.2	325.7
12.4	991.7	965.6	937.7	907.8	875.1	838.8	792.3	755.4	702.7	621.4	537.0	463.0	405.1	361.9
13.1	995.1	969.5	942.3	913.2	881.6	846.9	799.9	766.5	726.9	657.7	579.9	506.8	446.1	398.6
13.8	998.5	973.3	946.7	918.4	887.8	854.5	809.8	776.2	742.8	684.4	615.5	546.2	485.1	434.8
14.5	1001.7	976.9	950.9	923.3	893.7	861.7	821.3	785.8	752.9	703.4	644.2	580.8	521.3	469.7
15.2	1004.8	980.5	955.0	928.0	899.3	868.4	833.5	795.7	760.1	717.0	667.2	610.4	554.0	502.4
15.9	1007.9	983.9	958.9	932.5	904.6	874.8	845.1	805.7	766.3	727.5	685.6	635.5	582.8	532.4
16.5	1010.8	987.3	962.7	936.9	909.7	880.9	855.0	815.6	773.3	736.8	700.7	656.3	607.8	559.4
17.2	1013.8	990.5	966.4	941.1	914.5	886.6	862.7	824.9	782.0	746.2	713.6	673.6	629.0	583.2
17.9	1016.6	993.7	969.9	945.2	919.2	892.0	868.0	833.4	792.5	756.5	725.2	688.0	646.8	604.0
18.6	1019.4	996.8	973.4	949.1	923.6	897.1	871.2	840.8	804.4	767.7	736.3	700.4	662.0	622.0
19.3	1022.1	999.9	976.8	952.9	927.9	901.9	873.0	847.2	816.7	779.6	747.1	711.5	675.1	637.8
20.0	1024.7	1002.8	980.1	956.6	932.1	906.5	874.7	852.8	828.2	791.1	757.6	721.9	686.9	651.8
20.7	1027.3	1005.7	983.3	960.2	936.1	911.0	877.5	857.8	837.5	801.2	767.8	732.1	698.1	664.8
21.4	1029.9	1008.5	986.4	963.6	940.0	915.4	882.4	862.8	843.6	808.9	777.2	742.4	709.3	677.2
22.1	1032.4	1011.2	989.4	967.0	943.8	919.7	889.7	868.1	846.3	813.7	785.7	752.8	720.7	689.4
22.8	1034.8	1013.9	992.4	970.3	947.5	923.9	898.9	873.9	846.5	816.1	792.9	763.0	732.1	701.3
23.4	1037.2	1016.6	995.3	973.5	951.1	928.0	907.7	880.0	847.0	817.9	799.3	772.2	742.8	712.7
24.1	1039.6	1019.1	998.2	976.6	954.5	931.9	911.3	885.6	853.1	823.2	805.5	779.5	751.3	722.5
24.8	1041.9	1021.6	1001.0	979.7	957.9	935.4	902.2	888.9	873.2	838.9	813.4	783.1	755.1	728.8

Appendix 4

Percent difference (%) between calculated and actual CO2 density values
(100% * [(calculated - actual) / actual])

	Temperature (°C)													
Pressure (MPa)	-1.1	4.4	10.0	15.6	21.1	26.7	32.2	37.8	43.3	48.9	54.4	60.0	65.6	71.1
7.6	0.00	0.00	0.00	0.00	0.01	0.06	4.40	-7.20	-7.86	-0.64	2.85	2.72	1.41	0.41
8.3	0.00	0.00	0.00	-0.01	-0.01	-0.09	-5.92	13.75	7.51	-2.16	-4.16	-2.93	-1.19	-0.13
9.0	0.00	0.00	0.00	0.01	0.00	-0.02	-0.30	-4.90	7.83	2.26	-1.82	-2.31	-1.37	-0.46
9.7	0.00	0.00	0.00	0.00	0.00	0.02	1.87	-2.53	-2.27	3.11	1.62	0.09	-0.43	-0.32
10.3	0.00	-0.01	0.00	-0.01	0.00	0.04	1.95	-0.17	-4.69	-0.43	2.52	1.76	0.60	0.03
11.0	-0.01	0.00	0.00	0.00	0.01	0.03	1.08	0.87	-2.45	-2.48	0.94	1.83	1.10	0.35
11.7	0.00	0.01	0.00	0.00	0.00	0.01	0.06	0.98	-0.13	-1.83	-0.71	0.77	0.92	0.50
12.4	0.00	0.00	0.00	0.00	0.00	-0.01	-0.72	0.66	1.31	-0.44	-1.15	-0.43	0.24	0.36
13.1	-0.01	0.00	0.01	0.00	0.00	-0.02	-1.08	0.23	1.84	0.62	-0.75	-0.97	-0.40	0.05
13.8	0.00	0.00	0.00	0.00	-0.01	-0.03	-1.04	-0.15	1.64	1.03	-0.17	-0.91	-0.70	-0.20
14.5	0.00	0.00	0.00	0.00	0.00	-0.02	-0.69	-0.35	1.00	0.92	0.27	-0.48	-0.63	-0.34
15.2	0.00	0.00	-0.01	0.01	0.00	-0.01	-0.21	-0.38	0.23	0.47	0.46	-0.03	-0.34	-0.27
15.9	0.00	0.01	0.00	-0.01	0.00	0.01	0.27	-0.28	-0.46	-0.10	0.44	0.31	0.01	-0.09
16.5	0.01	0.01	0.00	0.00	0.00	0.02	0.62	-0.11	-0.91	-0.60	0.28	0.47	0.29	0.10
17.2	-0.01	0.00	0.00	0.00	0.00	0.01	0.73	0.05	-1.02	-0.89	0.08	0.45	0.42	0.25
17.9	0.01	-0.01	0.00	0.01	0.00	0.03	0.61	0.19	-0.80	-0.94	-0.08	0.31	0.40	0.31
18.6	-0.01	0.00	0.00	0.00	0.00	0.02	0.29	0.23	-0.35	-0.73	-0.16	0.11	0.23	0.26
19.3	-0.01	0.00	0.00	0.00	0.00	0.01	-0.16	0.21	0.22	-0.35	-0.16	-0.09	0.03	0.14
20.0	0.00	0.00	0.01	0.00	0.00	-0.01	-0.57	0.13	0.74	0.03	-0.05	-0.25	-0.19	0.01
20.7	0.01	0.00	0.00	0.00	0.00	-0.01	-0.85	0.02	1.03	0.29	0.06	-0.30	-0.32	-0.12
21.4	0.00	0.00	0.00	0.00	0.00	-0.01	-0.86	-0.05	0.99	0.33	0.19	-0.22	-0.29	-0.13
22.1	0.00	0.00	0.00	0.00	0.00	-0.01	-0.57	-0.07	0.56	0.05	0.24	-0.07	-0.16	-0.06
22.8	0.00	0.00	0.01	0.00	0.00	0.01	-0.06	-0.01	-0.12	-0.48	0.19	0.15	0.07	0.12
23.4	0.00	0.00	-0.01	0.00	0.00	0.03	0.43	0.12	-0.72	-1.03	0.10	0.29	0.30	0.28
24.1	0.00	0.00	0.01	0.00	0.00	0.03	0.35	0.21	-0.62	-1.09	0.02	0.23	0.28	0.32
24.8	0.00	0.00	0.00	0.00	0.00	-0.01	-1.10	0.05	1.10	0.08	0.20	-0.23	-0.28	-0.05

Appendix 5

CO2 Viscosity (Pa-s) as a Function of Temperature (^oC) and Pressure (MPa)
(actual values from Kinder Morgan)

	Temperature (^oC)													
Pressure (MPa)	-1.1	4.4	10.0	15.6	21.1	26.7	32.2	37.8	43.3	48.9	54.4	60.0	65.6	71.1
7.6	1.17E-04	1.07E-04	9.76E-05	8.86E-05	7.94E-05	6.92E-05	3.42E-05	2.26E-05	2.14E-05	2.15E-05	2.10E-05	2.07E-05	2.05E-05	2.05E-05
8.3	1.18E-04	1.08E-04	9.93E-05	9.04E-05	8.17E-05	7.24E-05	2.93E-05	2.68E-05	2.48E-05	2.33E-05	2.25E-05	2.19E-05	2.15E-05	2.13E-05
9.0	1.20E-04	1.10E-04	1.01E-04	9.22E-05	8.37E-05	7.51E-05	3.89E-05	3.35E-05	2.93E-05	2.62E-05	2.46E-05	2.35E-05	2.27E-05	2.22E-05
9.7	1.21E-04	1.11E-04	1.02E-04	9.39E-05	8.56E-05	7.74E-05	5.35E-05	4.35E-05	3.58E-05	3.02E-05	2.76E-05	2.58E-05	2.44E-05	2.35E-05
10.3	1.22E-04	1.13E-04	1.04E-04	9.55E-05	8.74E-05	7.95E-05	6.74E-05	5.34E-05	4.25E-05	3.47E-05	3.11E-05	2.86E-05	2.66E-05	2.53E-05
11.0	1.24E-04	1.14E-04	1.05E-04	9.71E-05	8.91E-05	8.14E-05	7.21E-05	5.80E-05	4.68E-05	3.87E-05	3.47E-05	3.17E-05	2.92E-05	2.74E-05
11.7	1.25E-04	1.16E-04	1.07E-04	9.85E-05	9.07E-05	8.31E-05	7.51E-05	6.15E-05	5.07E-05	4.25E-05	3.80E-05	3.44E-05	3.16E-05	2.93E-05
12.4	1.26E-04	1.17E-04	1.08E-04	1.00E-04	9.22E-05	8.48E-05	7.65E-05	6.41E-05	5.40E-05	4.61E-05	4.10E-05	3.70E-05	3.37E-05	3.12E-05
13.1	1.28E-04	1.18E-04	1.10E-04	1.01E-04	9.37E-05	8.64E-05	7.76E-05	6.63E-05	5.68E-05	4.92E-05	4.38E-05	3.94E-05	3.58E-05	3.30E-05
13.8	1.29E-04	1.20E-04	1.11E-04	1.03E-04	9.51E-05	8.79E-05	7.95E-05	6.85E-05	5.92E-05	5.17E-05	4.63E-05	4.18E-05	3.81E-05	3.50E-05
14.5	1.30E-04	1.21E-04	1.12E-04	1.04E-04	9.65E-05	8.94E-05	8.06E-05	7.04E-05	6.16E-05	5.43E-05	4.88E-05	4.42E-05	4.02E-05	3.70E-05
15.2	1.32E-04	1.22E-04	1.14E-04	1.05E-04	9.79E-05	9.08E-05	8.05E-05	7.18E-05	6.40E-05	5.72E-05	5.14E-05	4.65E-05	4.23E-05	3.88E-05
15.9	1.33E-04	1.23E-04	1.15E-04	1.07E-04	9.92E-05	9.22E-05	8.09E-05	7.33E-05	6.62E-05	5.98E-05	5.39E-05	4.87E-05	4.43E-05	4.07E-05
16.5	1.34E-04	1.25E-04	1.16E-04	1.08E-04	1.00E-04	9.35E-05	8.15E-05	7.47E-05	6.83E-05	6.21E-05	5.61E-05	5.08E-05	4.62E-05	4.25E-05
17.2	1.35E-04	1.26E-04	1.17E-04	1.09E-04	1.02E-04	9.48E-05	8.23E-05	7.62E-05	7.01E-05	6.42E-05	5.82E-05	5.28E-05	4.82E-05	4.43E-05
17.9	1.37E-04	1.27E-04	1.18E-04	1.10E-04	1.03E-04	9.61E-05	8.33E-05	7.76E-05	7.18E-05	6.61E-05	6.01E-05	5.47E-05	5.01E-05	4.61E-05
18.6	1.38E-04	1.28E-04	1.20E-04	1.12E-04	1.04E-04	9.73E-05	8.45E-05	7.89E-05	7.33E-05	6.78E-05	6.19E-05	5.66E-05	5.19E-05	4.79E-05
19.3	1.39E-04	1.30E-04	1.21E-04	1.13E-04	1.05E-04	9.85E-05	8.58E-05	8.02E-05	7.47E-05	6.92E-05	6.35E-05	5.83E-05	5.37E-05	4.97E-05
20.0	1.40E-04	1.31E-04	1.22E-04	1.14E-04	1.07E-04	9.97E-05	8.74E-05	8.15E-05	7.58E-05	7.04E-05	6.50E-05	6.00E-05	5.55E-05	5.15E-05
20.7	1.41E-04	1.32E-04	1.23E-04	1.15E-04	1.08E-04	1.01E-04	8.85E-05	8.28E-05	7.73E-05	7.19E-05	6.65E-05	6.16E-05	5.71E-05	5.31E-05
21.4	1.43E-04	1.33E-04	1.24E-04	1.16E-04	1.09E-04	1.02E-04	8.95E-05	8.41E-05	7.87E-05	7.34E-05	6.81E-05	6.31E-05	5.86E-05	5.46E-05
22.1	1.44E-04	1.34E-04	1.26E-04	1.17E-04	1.10E-04	1.03E-04	9.06E-05	8.53E-05	8.00E-05	7.48E-05	6.95E-05	6.45E-05	6.00E-05	5.60E-05
22.8	1.45E-04	1.35E-04	1.27E-04	1.19E-04	1.11E-04	1.04E-04	9.18E-05	8.65E-05	8.13E-05	7.61E-05	7.08E-05	6.59E-05	6.14E-05	5.74E-05
23.4	1.46E-04	1.37E-04	1.28E-04	1.20E-04	1.12E-04	1.05E-04	9.29E-05	8.77E-05	8.25E-05	7.73E-05	7.20E-05	6.71E-05	6.27E-05	5.87E-05
24.1	1.47E-04	1.38E-04	1.29E-04	1.21E-04	1.13E-04	1.06E-04	9.41E-05	8.88E-05	8.36E-05	7.85E-05	7.32E-05	6.83E-05	6.39E-05	5.99E-05
24.8	1.49E-04	1.39E-04	1.30E-04	1.22E-04	1.14E-04	1.08E-04	9.53E-05	8.99E-05	8.47E-05	7.95E-05	7.43E-05	6.94E-05	6.50E-05	6.11E-05

Appendix 6

CO2 Viscosity (Pa-s) as a Function of Temperature (^oC) and Pressure (MPa)
(calculated values from regression equations)

	Temperature (^oC)													
Pressure (MPa)	-1.1	4.4	10.0	15.6	21.1	26.7	32.2	37.8	43.3	48.9	54.4	60.0	65.6	71.1
7.6	1.17E-04	1.07E-04	9.76E-05	8.86E-05	7.94E-05	6.92E-05	3.05E-05	2.13E-05	2.08E-05	2.12E-05	2.10E-05	2.07E-05	2.06E-05	2.05E-05
8.3	1.18E-04	1.08E-04	9.93E-05	9.04E-05	8.16E-05	7.23E-05	3.45E-05	2.80E-05	2.52E-05	2.34E-05	2.24E-05	2.17E-05	2.13E-05	2.11E-05
9.0	1.20E-04	1.10E-04	1.01E-04	9.22E-05	8.37E-05	7.50E-05	4.27E-05	3.58E-05	3.04E-05	2.66E-05	2.48E-05	2.36E-05	2.27E-05	2.22E-05
9.7	1.21E-04	1.11E-04	1.02E-04	9.39E-05	8.56E-05	7.74E-05	5.23E-05	4.35E-05	3.59E-05	3.04E-05	2.78E-05	2.60E-05	2.46E-05	2.37E-05
10.3	1.22E-04	1.13E-04	1.04E-04	9.55E-05	8.74E-05	7.95E-05	6.15E-05	5.05E-05	4.12E-05	3.44E-05	3.11E-05	2.86E-05	2.68E-05	2.54E-05
11.0	1.24E-04	1.14E-04	1.05E-04	9.71E-05	8.91E-05	8.14E-05	6.92E-05	5.65E-05	4.61E-05	3.84E-05	3.44E-05	3.14E-05	2.91E-05	2.73E-05
11.7	1.25E-04	1.16E-04	1.07E-04	9.85E-05	9.07E-05	8.32E-05	7.49E-05	6.13E-05	5.04E-05	4.22E-05	3.77E-05	3.42E-05	3.14E-05	2.92E-05
12.4	1.26E-04	1.17E-04	1.08E-04	1.00E-04	9.22E-05	8.48E-05	7.87E-05	6.49E-05	5.41E-05	4.58E-05	4.08E-05	3.69E-05	3.37E-05	3.12E-05
13.1	1.28E-04	1.18E-04	1.10E-04	1.01E-04	9.37E-05	8.64E-05	8.07E-05	6.76E-05	5.72E-05	4.92E-05	4.38E-05	3.95E-05	3.59E-05	3.31E-05
13.8	1.29E-04	1.20E-04	1.11E-04	1.03E-04	9.51E-05	8.79E-05	8.14E-05	6.95E-05	5.99E-05	5.22E-05	4.66E-05	4.19E-05	3.81E-05	3.50E-05
14.5	1.30E-04	1.21E-04	1.12E-04	1.04E-04	9.65E-05	8.94E-05	8.12E-05	7.09E-05	6.22E-05	5.49E-05	4.92E-05	4.43E-05	4.02E-05	3.69E-05
15.2	1.32E-04	1.22E-04	1.13E-04	1.05E-04	9.79E-05	9.08E-05	8.07E-05	7.20E-05	6.43E-05	5.74E-05	5.16E-05	4.65E-05	4.23E-05	3.88E-05
15.9	1.33E-04	1.23E-04	1.15E-04	1.07E-04	9.92E-05	9.22E-05	8.02E-05	7.30E-05	6.61E-05	5.97E-05	5.38E-05	4.86E-05	4.43E-05	4.07E-05
16.5	1.34E-04	1.25E-04	1.16E-04	1.08E-04	1.00E-04	9.35E-05	8.02E-05	7.41E-05	6.79E-05	6.19E-05	5.60E-05	5.07E-05	4.62E-05	4.25E-05
17.2	1.35E-04	1.26E-04	1.17E-04	1.09E-04	1.02E-04	9.48E-05	8.08E-05	7.54E-05	6.96E-05	6.39E-05	5.80E-05	5.27E-05	4.82E-05	4.44E-05
17.9	1.37E-04	1.27E-04	1.18E-04	1.10E-04	1.03E-04	9.61E-05	8.21E-05	7.69E-05	7.13E-05	6.57E-05	5.99E-05	5.46E-05	5.01E-05	4.62E-05
18.6	1.38E-04	1.28E-04	1.20E-04	1.12E-04	1.04E-04	9.73E-05	8.38E-05	7.86E-05	7.30E-05	6.75E-05	6.18E-05	5.65E-05	5.19E-05	4.80E-05
19.3	1.39E-04	1.30E-04	1.21E-04	1.13E-04	1.05E-04	9.85E-05	8.59E-05	8.03E-05	7.47E-05	6.92E-05	6.35E-05	5.83E-05	5.37E-05	4.97E-05
20.0	1.40E-04	1.31E-04	1.22E-04	1.14E-04	1.07E-04	9.97E-05	8.79E-05	8.20E-05	7.63E-05	7.07E-05	6.52E-05	6.00E-05	5.55E-05	5.14E-05
20.7	1.41E-04	1.32E-04	1.23E-04	1.15E-04	1.08E-04	1.01E-04	8.96E-05	8.35E-05	7.77E-05	7.22E-05	6.67E-05	6.16E-05	5.71E-05	5.30E-05
21.4	1.43E-04	1.33E-04	1.24E-04	1.16E-04	1.09E-04	1.02E-04	9.07E-05	8.47E-05	7.90E-05	7.36E-05	6.82E-05	6.31E-05	5.87E-05	5.46E-05
22.1	1.44E-04	1.34E-04	1.26E-04	1.17E-04	1.10E-04	1.03E-04	9.12E-05	8.57E-05	8.02E-05	7.48E-05	6.95E-05	6.45E-05	6.01E-05	5.60E-05
22.8	1.45E-04	1.35E-04	1.27E-04	1.19E-04	1.11E-04	1.04E-04	9.11E-05	8.63E-05	8.11E-05	7.60E-05	7.08E-05	6.58E-05	6.14E-05	5.74E-05
23.4	1.46E-04	1.37E-04	1.28E-04	1.20E-04	1.12E-04	1.05E-04	9.11E-05	8.70E-05	8.21E-05	7.71E-05	7.19E-05	6.70E-05	6.27E-05	5.86E-05
24.1	1.47E-04	1.38E-04	1.29E-04	1.21E-04	1.13E-04	1.06E-04	9.22E-05	8.82E-05	8.33E-05	7.83E-05	7.31E-05	6.81E-05	6.39E-05	5.99E-05
24.8	1.49E-04	1.39E-04	1.30E-04	1.22E-04	1.14E-04	1.08E-04	9.63E-05	9.07E-05	8.51E-05	7.97E-05	7.44E-05	6.94E-05	6.51E-05	6.11E-05

Percent difference (%) between calculated and actual CO_2 viscosity values
(100% * [(calculated - actual) / actual])

	Temperature (°C)													
Pressure (MPa)	-1.1	4.4	10.0	15.6	21.1	26.7	32.2	37.8	43.3	48.9	54.4	60.0	65.6	71.1
7.6	0.00	0.00	0.00	0.00	0.01	0.05	-10.75	-5.75	-2.75	-1.08	-0.33	0.09	0.38	0.47
8.3	0.00	0.00	0.00	0.00	-0.01	-0.09	18.05	4.83	1.82	0.53	-0.33	-0.83	-1.00	-0.95
9.0	0.00	0.00	0.00	0.00	-0.01	-0.03	9.81	6.85	4.03	1.69	0.98	0.50	0.19	-0.07
9.7	0.00	0.00	0.00	-0.01	0.00	0.02	-2.16	0.05	0.49	0.64	0.78	0.82	0.79	0.59
10.3	0.00	0.00	0.00	0.00	0.01	0.04	-8.71	-5.27	-2.94	-0.91	-0.17	0.17	0.47	0.52
11.0	0.00	0.00	0.00	0.00	0.00	0.03	-4.06	-2.51	-1.58	-0.85	-0.74	-0.72	-0.55	-0.36
11.7	0.00	0.00	0.00	0.00	0.01	0.02	-0.20	-0.41	-0.57	-0.70	-0.70	-0.70	-0.52	-0.35
12.4	0.00	0.00	0.00	0.00	0.00	-0.01	2.88	1.26	0.24	-0.49	-0.45	-0.35	-0.14	-0.03
13.1	0.00	0.00	0.00	0.00	0.00	-0.02	4.04	1.93	0.71	-0.10	0.04	0.22	0.42	0.40
13.8	0.00	0.00	0.00	0.00	0.00	-0.03	2.33	1.41	1.14	0.96	0.67	0.37	0.18	-0.04
14.5	0.00	0.00	0.00	0.00	0.00	-0.02	0.67	0.72	1.05	1.22	0.81	0.29	-0.03	-0.30
15.2	0.00	0.00	0.00	0.00	0.00	-0.01	0.22	0.27	0.36	0.39	0.29	0.09	0.05	-0.06
15.9	0.00	0.00	0.00	0.00	0.00	0.00	-0.81	-0.35	-0.18	-0.06	-0.02	-0.08	0.03	0.00
16.5	0.00	0.00	0.00	0.00	0.00	0.01	-1.56	-0.80	-0.57	-0.40	-0.24	-0.19	0.02	0.07
17.2	0.00	0.00	0.00	0.00	0.01	0.02	-1.79	-0.96	-0.73	-0.57	-0.36	-0.24	0.02	0.10
17.9	0.00	0.00	0.00	0.00	0.01	0.01	-1.48	-0.81	-0.67	-0.57	-0.35	-0.24	0.01	0.09
18.6	0.00	0.00	0.00	0.00	0.00	0.01	-0.77	-0.42	-0.39	-0.36	-0.23	-0.20	0.03	0.06
19.3	0.00	0.00	0.00	0.00	0.00	0.00	0.03	0.10	0.05	0.01	0.01	-0.11	0.00	-0.03
20.0	0.00	0.00	0.00	0.00	0.00	-0.01	0.58	0.58	0.59	0.57	0.35	-0.01	-0.06	-0.21
20.7	0.00	0.00	0.00	0.00	0.00	-0.02	1.31	0.85	0.60	0.42	0.29	0.04	0.05	-0.08
21.4	0.00	0.00	0.00	0.00	0.00	-0.02	1.35	0.80	0.47	0.24	0.19	0.02	0.13	0.02
22.1	0.00	0.00	0.00	0.00	0.00	-0.01	0.60	0.43	0.19	0.04	0.07	-0.04	0.12	0.06
22.8	0.00	0.00	0.00	0.00	0.00	0.00	-0.72	-0.16	-0.15	-0.15	-0.06	-0.14	0.08	0.04
23.4	0.00	0.00	0.00	0.00	0.00	0.01	-1.99	-0.71	-0.41	-0.25	-0.15	-0.23	0.00	-0.02
24.1	0.00	0.00	0.00	0.00	0.01	0.01	-2.03	-0.65	-0.33	-0.19	-0.10	-0.25	-0.03	-0.06
24.8	0.00	0.00	0.00	0.00	-0.01	-0.02	1.05	0.85	0.50	0.26	0.20	-0.03	0.12	-0.01

Appendix 8

<u>SECTION III:</u>
Comparing Techno-Economic Models for Pipeline Transport of Carbon Dioxide

David L. McCollum
dlmccollum@ucdavis.edu

Institute of Transportation Studies
University of California
One Shields Avenue
Davis, CA 95616

ABSTRACT

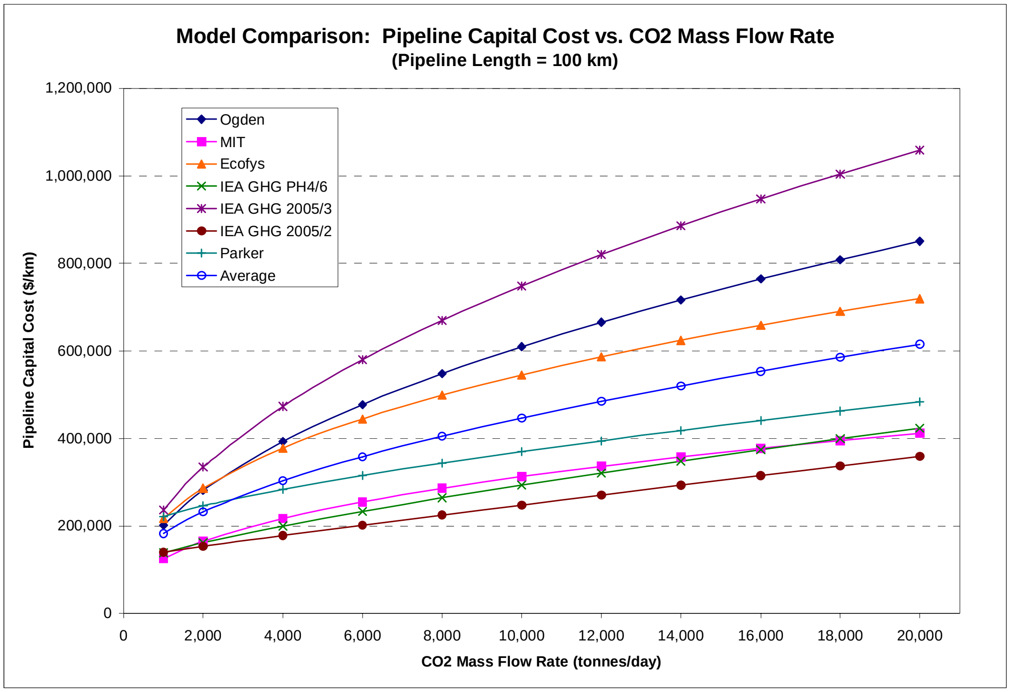
Due to a heightened interest in technologies to mitigate global climate change, research in the field of carbon capture and storage (CCS) has increased in recent years, with the goal of answering the many questions that still remain in this uncertain field. At the top of the list of key issues are CCS costs: costs of carbon dioxide (CO2) capture, compression, transport, storage, and so on. This paper focuses on costs of CO2 pipeline transport. Several recent techno-economic models for estimating pipeline sizes and costs are compared on an "apples-to-apples" basis by applying the same set of input assumptions across all models. We find that there is a large degree of variability between the output of the different models, particularly among the cost estimates, that stems from the differing approaches that each model employs. By averaging the cost estimates of the models over a wide range of CO2 mass flow rates and pipeline lengths, we have created a new CO2 pipeline capital cost model that is a function only of CO2 mass flow rate and pipeline length. This removes the need to calculate the pipeline diameter in advance of calculating costs. We feel that this equation is a reliable estimator of mid-range costs, given that it has been derived from a number of recent, reliable studies on CCS.

Keywords: carbon dioxide, CO2, CO_2, CCS, pipeline, transport, sequestration, techno-economic, cost model, climate change, greenhouse gas

EXECUTIVE SUMMARY

Due to a heightened interest in technologies to mitigate global climate change, research in the field of carbon capture and storage (CCS) has increased in recent years, with the goal of answering the many questions that still remain in this uncertain field. At the top of the list of key questions are CCS costs: costs of carbon dioxide (CO2) capture, costs of transport, costs of storage, and so on. Although the practice of transporting and storing CO2 underground has been around for a few decades, as it is used in the oil and gas industry for enhanced oil recovery (EOR), predicting the economics is still uncertain. In light of this, several studies have developed CCS models to try and predict costs, particularly for transport and storage. These models, however, differ in many ways, namely in their cost and flow equations, assumptions for operating conditions, and the reference years that their costs are expressed in. Thus, the models' output—e.g., pipeline diameter, capital cost, O&M costs, levelized CO2 costs, etc.—comes out differently, making it difficult to compare the models' predicted costs on an "apples-to-apples" basis. By replicating the models and applying some of the same key assumptions across all models, comparisons can be made, similarities/differences can be noted, and new models can be generated that are essentially a combination of all models. We have carried out this procedure for a few of the more recent CO2 transport models. The scope of this study was limited to onshore pipelines, since they are likely to be the most cost-effective and realistic means for transporting CO2 in the future, at least in the United States. The transport models that were compared came from the following studies: Ogden, MIT, Ecofys, IEA GHG PH4/6, IEA GHG 2005/2, IEA GHG 2005/3, and Parker. Each of these studies was carried out within the last four years; and except for the models of Parker, which use natural gas pipeline costs to predict the costs of hydrogen pipelines, all of the models are geared specifically towards CO2 pipelines. In this paper, the basic concepts, equations, and assumptions of the above models are discussed; though, the reader is encouraged to consult the original reports for a more thorough description. The key similarities and differences between the models are then highlighted. And ultimately, a set of common basis assumptions is decided upon, with new models being created that are essentially a combination of all seven of the original models.

The wide variability in the costs that each of the models estimates is easily seen in the following graph:



By averaging the estimated capital costs of all models over a range of flow rates and pipeline lengths, we have created the following equation to model pipeline capital cost:

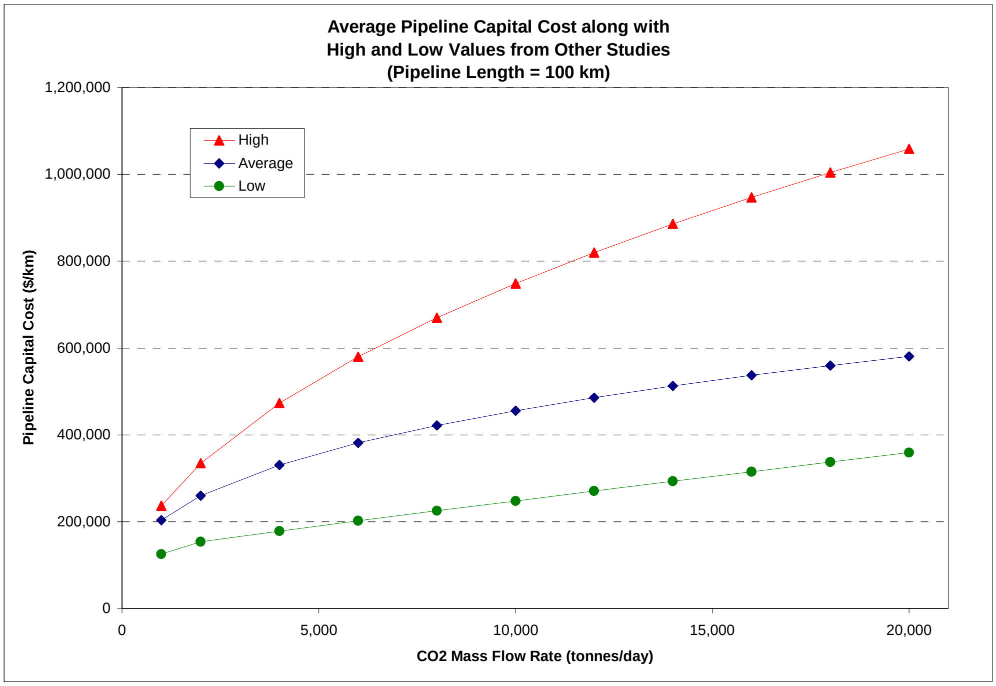
Pipeline Capital Cost [$/km] = $(9970 * m^{0.35}) * L^{0.13}$
(where m = CO2 mass flow rate [tonnes/day], and L = pipeline length [km])

Costs are given in year 2005 US dollars. This equation provides a method of estimating the pipeline capital cost per unit length based on two quantities that are typically known—CO2 mass flow rate and pipeline length. By approaching the capital cost in this way, one can avoid the calculation of pipeline diameter in advance, which can be advantageous. The above equation is a reliable method for calculating pipeline capital cost since it is essentially derived from seven other pipeline models, all of which are recent and reliable. The upper and lower bounds for the pipeline capital cost are found to be given by the following equations:

Pipeline Capital Cost (Low) [$/km] = $(8500 * m^{0.35}) * L^{0.06}$

Pipeline Capital Cost (High) [$/km] = $(4100 * m^{0.50}) * L^{0.13}$
(where m = CO2 mass flow rate [tonnes/day], and L = pipeline length [km])

The estimates for average pipeline capital cost, along with the upper and lower bounds are shown together in the following graph.



One can feel comfortable in knowing that while the uncertainty of CO_2 pipeline capital cost may be great, it will very likely be within these upper and lower bounds, and probably close to the average.

INTRODUCTION

Due to a heightened interest in technologies to mitigate global climate change, research in the field of carbon capture and storage (CCS) has increased in recent years, with the goal of answering the many questions that still remain in this uncertain field. At the top of the list of key questions are CCS costs: costs of carbon dioxide (CO2) capture, costs of transport, costs of storage, and so on. Although the practice of transporting and storing CO2 underground has been around for a few decades, as it is used in the oil and gas industry for enhanced oil recovery (EOR), predicting the economics is still uncertain. In light of this, several studies have developed CCS models to try and predict costs, particularly for transport and storage. These models, however, differ in many ways, namely in their cost and flow equations, assumptions for operating conditions, and the reference years that their costs are expressed in. Thus, the models' output—e.g., pipeline diameter, capital cost, O&M costs, levelized CO2 costs, etc.—comes out differently, making it difficult to compare the models' predicted costs on an "apples-to-apples" basis. By replicating the models and applying some of the same key assumptions across all models, comparisons can be made, similarities/differences can be noted, and new models can be generated that are essentially a combination of all models. We have carried out this procedure for a few of the more recent CO2 transport models. The scope of this study was limited to onshore pipelines, since they are likely to be the most cost-effective and realistic means for transporting CO2 in the future, at least in the United States. The transport models that were compared came from the following studies: Ogden [1], MIT [2], Ecofys [3], IEA GHG PH4/6 [4], IEA GHG 2005/2 [5], IEA GHG 2005/3 [6], and Parker [7]. Each of these studies was carried out within the last four years; and except for the models of Parker, which use natural gas pipeline costs to predict the costs of hydrogen pipelines, all of the models are geared specifically towards CO2 pipelines. In this paper, the basic concepts, equations, and assumptions of the above models are discussed; though, the reader is encouraged to consult the original reports for a more thorough description. The key similarities and differences between the models are then highlighted. And ultimately, a set of common basis assumptions is decided upon, with new models being created that are essentially a combination of all seven of the original models.

DESCRIPTION OF MODELS

The Ogden Models

The CO2 transport models used in Ogden's report (or rather, those described in detail in Appendix C of the full report) were created to model a hydrogen production and distribution infrastructure that makes use of CCS. Although the publication date of the report is 2004 (i.e., later than some of the other models that will also be described here), work on the report began much earlier in 2002 and for this reason did not build upon models that have come out more recently.

For starters, Ogden's models use a complex equation for calculating the volumetric flow rate (Q) of CO2, which was adapted from Farris [8]:

$$Q = C_1 \sqrt{(1/f)}\ [(P_{inpipe}^2 - P_{outpipe}^2 - C_2\{G\Delta hP^2_{avg} / Z_{avg}\ T_{avg}\}) / (G\ T_{avg}\ Z_{avg}\ L)]^{0.5}\ D^{2.5}\ E$$

(where Q = CO2 flow rate [Nm^3/s], C_1 = 18.921, f = friction factor, P_{inpipe} = pipeline inlet pressure [kPa], $P_{outpipe}$ = pipeline outlet pressure [kPa], C_2 = 0.06836, G = CO2 specific gravity = 1.519, Δh = change in elevation [m], P_{avg} = average pipeline pressure, Z_{avg} = CO2 compressibility at P_{avg}, T_{avg} = average temperature [K], L = pipeline length [km], D = pipeline diameter [m], E = pipeline efficiency)

Oftentimes, however, one already knows the CO2 mass flow rate (e.g., in tonnes/day), which can be converted to volumetric flow rate, thus enabling the back-calculation of pipeline diameter. When using the above equation to solve for diameter, we assumed that some of the variables had the following constant values: change in elevation, Δh=0; CO2 compressibility, Z_{avg}=0.25; pipeline efficiency, E=1.0.

The calculated diameter seems to be sensitive to compressibility and, especially, efficiency. We have done a simple sensitivity analysis for both at a given set of operating conditions. Ogden suggests that compressibility will be in the range of 0.17-0.30 for pure CO2 at average pipeline pressures of 8.8-12.0 MPa and temperatures from less than 20 ^oC up to 40 ^oC. By our calculations, at a representative CO2 mass flow rate of 10,000 tonnes/day, inlet and outlet pressures of 15.2 and 10.3 MPa, respectively, and a temperature of 25 ^oC, a 76% increase in compressibility (from 0.17 to 0.30) will lead to a calculated pipeline diameter increase of only 11% (from 10.6 to 11.8 inches). Thus, when replicating Ogden's models, we assume that Z_{avg}=0.25, which according to Ogden is a reasonable estimate at the temperature and average pipeline pressure that we will generally consider—25 ^oC and 12.75 MPa, respectively. Similarly, we calculate that, at the same operating conditions mentioned above, a decrease in pipeline efficiency of 25 percentage points (from 100% to 75%) will lead to a calculated pipeline diameter increase of 11% (from 11.4 to 12.7 inches). The dependence of calculated diameter on pipeline efficiency gets much stronger, however, as efficiency gets lower and lower; for example, the next 25% percentage point decrease in pipeline efficiency (from 75% to 50%) will lead to a calculated pipeline diameter increase of 17% (from 12.7 to 14.9 inches). Ogden does not suggest any values for pipeline efficiency in her report, so we simply assume E=1.0 when replicating her models; this assumption may not reflect real world pipelines but seems reasonable here, since none of the other models under study in this report consider pipeline efficiency. Thus, we are essentially canceling out the effect that pipeline efficiency may have so that the pipeline diameter calculated by Ogden's models can be more directly compared to that calculated from other models.

Notice also that Ogden calculates the friction factor (f) by the Nikuradse equation; this factor is similar in magnitude to the Fanning friction factor in that it is four times smaller than friction factors used in some of the other models described later in this report. Yet, even after accounting for the factor of four difference, the Nikuradse equation calculates friction factors that are consistently smaller than those assumed in the other models, which affects the size of the calculated pipeline diameter, as will be shown later. Furthermore, a word of caution that might be helpful to others when using the above-mentioned flow rate equation is to multiply the length term, L, in the equation by 1000. Although not explicitly stated in the description of the equation, this causes the units to work out and helps calculate a flow rate (or alternately, a pipeline diameter) that is on the correct order of magnitude.

To estimate pipeline capital costs, the Ogden models use capital cost estimates from Skovholt's 1993 study [9]. These estimates give capital costs (in \$/m) for four different sizes of pipeline diameter (16, 30, 40, and 64 inches). With these four data points, an equation is generated that scales up the capital cost as the diameter gets larger. And finally, the capital cost (in \$/m) is multiplied by the pipeline length (L) to calculate the total capital cost. These two equations are shown below:

$$\text{Capital Cost (\$/m)} = \$700/\text{m} \times (D / 16\ \text{in})^{1.2}$$

(where D = pipeline diameter [inches])

$$\text{Total Capital Cost (\$)} = \text{Capital Cost (\$/m)} \times L\ (\text{m})$$

(where L = pipeline length [m])

Ogden prefers, however, to use capital cost equations that are functions directly of CO2 flow rate (Q) and pipeline length (L), rather than diameter (D), thus making it possible to calculate the pipeline capital costs without having to solve for D directly. However, the cost equations are indirectly functions of diameter, since it has simply been parameterized away using other variables. The equation is:

$$\text{Capital Cost (\$/m)} = \$700/\text{m} \times (Q / 16{,}000\ \text{tonnes/day})^{0.48} \times (L / 100\ \text{km})^{0.24}$$

(where Q = CO2 mass flow rate [tonnes/day], and L = pipeline length [kilometers])

$$\text{Total Capital Cost (\$)} = \text{Capital Cost (\$/m)} \times L\ (\text{m})$$

When replicating and comparing models in this report, we use Ogden's latter capital cost models, making them the only models (of those that are compared) that do not use pipeline diameter to calculate costs.

Ogden uses the following equation to calculate the levelized cost of CO2 transport:

$$\text{Levelized Cost (\$/tonne CO2)} = (\text{CRF} + \text{O\&M}) \times \text{Total Capital Cost} / [Q\ (\text{Nm}^3/\text{s}) \times 3.17 \times 10^7\ \text{sec/year} \times (1.965\ \text{kg CO}^2/\text{Nm}^3) / (1000\ \text{kg/tonne})]$$

(where CRF = capital recovery factor = 0.15, and O&M = O&M cost factor = 0.04)

Finally, note that all of Ogden's costs are expressed in year 2001 US\$.

The MIT Models

The Massachusetts Institute of Technology's Laboratory for Energy and the Environment published a study on the economics of CO2 storage in 2003. Chapter 2 of their report outlines a methodology for calculating CO2 pipeline diameter and costs; this process is iterative. First, one has to guess a value for the pipeline diameter (D). Second, the Reynold's number (Re) is calculated by the following equation:

$$Re = 4\ m / (\pi\ \mu\ D)$$

(where m = CO2 mass flow rate, D = pipeline diameter, and μ = CO2 viscosity)

With the calculated Reynold's number and the MIT study's assumed pipeline roughness factor (ε) of 0.00015 feet, the Fanning friction factor (f) is found by using a Moody chart. This method, however, would require a manual look-up for each iteration, so MIT uses an empirical relation based on the Moody chart [10].

$$f = \frac{1}{4\left[-1.8\log_{10}\left\{\frac{6.91}{Re} + \left(\frac{12(\varepsilon / D)}{3.7}\right)^{1.11}\right\}\right]^2}$$

(f = friction factor, Re = Reynold's number, ε = roughness factor [ft], and D = pipeline diameter [in])

Next, the diameter is calculated by the following equation:

$$D^5 = (32\ f\ m^2)/\ (\pi^2\ \rho\ (\Delta P/\Delta L))$$

(where ΔP = inlet pressure – outlet pressure, and ΔL = pipeline length)

The diameter calculated by this equation is then compared to the previously guessed value of diameter. If the calculated diameter is much different from the guessed value, then the calculated value is used to re-calculate a new Reynold's number, friction factor, and diameter. This process is repeated until the calculated diameter is the same as the one used at the start of the iteration. It should also be noted that MIT assumes a CO2 density and viscosity of 884 kg/m^3 and 6.06 x 10^{-5} N-s/m^2, respectively, for their pipeline transport models.

To calculate CO2 pipeline costs, the MIT study uses historical cost data for natural gas pipeline construction, as reported in the *Oil & Gas Journal*. From this data, they conclude that, on average, construction costs for CO2 pipelines would be $20,989/in/km. Furthermore, based on estimates by Fox [11], they suggest that O&M costs, other than pumping, would be $3,100/km/year, independent of pipeline diameter. Thus, the total annual cost and levelized cost are calculated by the following equations:

Total Annual Cost ($/yr) = {($20,989/in/km) x D x L x CRF} + {($3,100/km/yr) x L}

(where D = pipeline diameter [in], L = pipeline length [km], and CRF = Capital Recovery Factor = 0.15/yr)

Levelized Cost ($/tonne CO2) = Total Annual Cost ($/yr) / { m x CF x 365 }

(where m = CO2 mass flow rate [tonnes/day], CF = Plant Capacity Factor = 0.80, and 365 = days per year)

Finally, since the MIT study does not state the reference year that they express costs in, we assume that they use year 1998 dollars, owing to the fact that 1998 is the most recent year for which they obtained natural gas pipeline cost data from the *Oil & Gas Journal*.

The Ecofys Models

The Ecofys models for CO_2 transport are part of a larger report to the European Commision on the potential of CCS as a cost-effective strategy to meet Kyoto Protocol targets for emissions reduction in the European Union. The technical and cost aspects of CO_2 transport are given in Appendix 3 of the report.

First, the equation for back-calculating the pipeline diameter is:

$$\Delta P = \lambda * (L/D) * (1/2) * \rho * v^2$$

(where ΔP = pressure drop [Pa], λ = friction factor, L = pipeline length [m], D = pipeline diameter [m], ρ = CO_2 density [kg/m^3], v = average flow velocity [m/s])

In the above flow equation, the velocity term, *v*, is a function of the mass flow rate and the cross-sectional area (i.e., diameter) of the pipeline. Thus, the equation can be rearranged to form the following equation:

$$D^5 = (8\ \lambda\ m^2)/ (\pi^2\ \rho\ (\Delta P/L))$$

(where m = CO_2 mass flow rate)

Note how similar this equation is to the analogous diameter equation that MIT uses. There are only two main differences: (1) The lead constant in the Ecofys equation is 8 versus 32 in the MIT equation—four times smaller because (2) The friction factor in the Ecofys equation (λ) is four times larger than the Fanning friction factor (f) in the MIT equation. In other words, the Ecofys and MIT equations for calculating pipeline diameter are essentially the same. The only other difference is that the Ecofys study assumes a constant friction factor, whereas MIT uses an equation to calculate the friction factor, as it is a function of Reynold's number. Ecofys suggests that their friction factor would be less than 1.5×10^{-2} for perfectly smooth pipeline walls and 2.0×10^{-2} for new untreated steel. (Actually, in the report Ecofys states that the friction factor for new untreated steel would be 2.0×10^{2}, i.e. the negative sign in the exponent is missing. Though, we believe this to be a typographical error, since using a friction factor of this magnitude would lead to an unusually large pipeline diameter.)

The equation used to calculate total pipeline capital cost is given by:

$$\text{Total Capital Cost (€)} = (1100\ \text{€/m}^2) * F_T * D * L$$

(where F_T = correction factor for terrain = 1 for most common terrain,
D = pipeline diameter [m], L = pipeline length [m])

The total capital cost is annualized with a 10% discount rate over a 25 year operational lifetime by the following equation:

$$Annual\ Capital\ Cost\ (euros/yr) = \frac{Total\ Capital\ Cost\ (euros)}{\dfrac{(1+i)^n - 1}{i(1+i)^n}}$$

(where n = operational lifetime [years], and i = discount rate)

The annual O&M costs are calculated as 2.1% of the total capital cost. And the total annual cost is found by summing the annual capital and O&M costs.

Annual O&M Costs (€/yr) = (O&M factor) * Total Capital Cost
(where O&M factor = 2.1%)

Total Annual Cost (€/yr) = Annual Capital Cost + Annual O&M Costs

The Ecofys study does not discuss the method used for calculating levelized cost of CO2 transport, but one can assume that it is similar to that used in other studies, for example, the MIT study's equation, which is shown below.

Levelized Cost (€/tonne CO2) = Total Annual Cost ($/yr) / { m * CF * 365 }
(where m = CO2 mass flow rate [tonnes/day], CF = Plant Capacity Factor, and 365 = days per year)

Finally, since the Ecofys study does not state the reference year that they express costs in, we assume that they use year 2003 euros since that is the year that their report was published.

The IEA GHG PH4/6 Models

In 2002, Woodhill Engineering Consultants of the United Kingdom studied the transmission of CO2 and energy for the IEA Greenhouse Gas R&D Programme. They wrote a report on the subject, as well as created a spreadsheet-based computer model for estimating the costs and performance of CO2 transport.

To calculate CO2 pipeline diameter, the IEA GHG PH4/6 study uses the following equation [12]:

$$\Delta P = 2.252 \frac{f \; L \; \rho \; Q^2}{D^5}$$

(where ΔP = pressure drop [bar], f = friction factor, L = pipeline length [km], ρ = CO2 density [kg/m^3], Q = CO2 flow rate [liter/min], and D = pipeline internal diameter [mm])

A friction factor (f) of 0.015 is assumed in the model. Further, the report states that a friction factor of this value is "relatively conservative in that it is likely to slightly oversize a liquid line rather than undersize it" [4, p. 3.26]. (Note that this friction factor is four times larger than the Fanning friction factor used in other studies.) With the internal diameter of the pipeline, the spreadsheet model uses a look-up table to find the closest nominal pipe size. We, however, did not have access to the look-up table, so when replicating the models of the IEA GHG PH4/6 study, we simply use the internal pipeline diameter throughout (e.g., in the pipeline cost calculations).

Woodhill Engineering developed several pipeline cost equations for the IEA GHG PH4/6 study based on in-house estimates. For onshore pipelines, they give three equations, one for each of three different ANSI piping classes: 600# ($P \leq 90$ bar), 900# ($P \leq 140$ bar), and 1500# ($P \leq 225$ bar). At the higher pressures likely required for CO2 transport, the ANSI Class 1500# pipe would be used. The capital cost equation for ANSI Class 1500# pipe is given as:

$$\text{Pipeline Capital Cost (\$)} = F_L * F_T * 10^6 * [\, (0.057 * L + 1.8663) + (0.00129 * L) * D + (0.000486 * L + 0.000007) * D^2 \,]$$

(where F_L = location factor, F_T = terrain factor, L = pipeline length [km], and D = pipeline diameter [in])

Location factors (F_L) for a few world regions are reproduced here: USA/Canada=1.0, Europe=1.0, UK=1.2, Japan=1.0, Australia=1.0. (A full list of location factors for all world regions can be found in the original IEA GHG PH4/6 report.) Terrain factors (F_T) are as follows: cultivated land=1.10, grassland=1.00, wooded=1.05, jungle=1.10, stony desert=1.10, <20% mountainous=1.30, >50% mountainous=1.50.

Booster stations for raising the CO2 pressure during pipeline transport are also considered in the IEA GHG PH4/6 models. In fact, the user of the spreadsheet model has the choice of whether or not to include booster stations. If booster stations are included, their capital costs can be calculated by the following equation:

$$\text{Booster Station Capital Cost (\$)} = N_B * F_L * (7.82 * \text{Power} + 0.46) * \$1{,}000{,}000$$

(where N_B = number of booster stations, F_L = location factor, and Power = pump power [MW])

'Power' is calculated by the following equation given in [13]:

$$\text{Power (MW)} = (Q * \Delta P) / (36{,}000 * \eta)$$

(where Q = CO2 flow rate [m^3/hr], ΔP = pressure increase through booster [bar], η = pump efficiency = 0.75)

The total capital cost is given by:

Total Capital Cost ($) = Pipeline Capital Cost + Booster Station Capital Cost

Equations for O&M costs were also developed for the IEA GHG PH4/6 study. The O&M cost equation for liquid CO2 onshore pipelines is given by:

Annual Pipeline O&M Costs ($/yr) = 120,000 + 0.61(23,213 * D + 899 * L – 259,269)
+ 0.7(39,305 * D + 1694 * L – 351,355) + 24,000
(where D = pipeline diameter [in], and L = pipeline length [km])

Similarly, booster station O&M costs (both fixed and variable) are also calculated. For fixed O&M costs, a look-up table is used. This table provides fixed O&M costs as a function of pump power (from 0 to 2 MW). To avoid the look-up table, we have created a second order regression equation that fits the fixed O&M cost vs. pump power with an R^2 value of 0.93. (Be advised that this equation should only be used in the range of 0-2 MW, since the second order equation is parabolic and will eventually begin predicting increasingly lower costs as the pump power increases.) The equation for booster station fixed O&M costs is given below:

Annual Booster Station Fixed O&M Costs ($/yr) = N_B * [-179,864 * $Power^2$ + 671,665 * Power + 159,292]
(where N_B = number of booster stations, and Power = pump power [MW])

The booster station variable O&M costs are calculated by the following equation:

Booster Station Variable O&M Costs ($/yr) = N_B * COE * Power * CF * (1000 kW/MW) * (24 hr/day) * (365 days/yr)
(where N_B = number of booster stations, COE = cost of electricity [$/kWh],
Power = pump power [MW], CF = plant capacity factor)

The total annual O&M costs are then:

Total Annual O&M Costs ($/yr) = Annual Pipeline O&M Costs
+ Annual Booster Station Fixed O&M Costs
+ Annual Booster Station Variable O&M Costs

And finally, the total annual cost and levelized cost are calculated by the following equations:

Total Annual Cost ($/yr) = (Total Capital Cost * CRF) + Total Annual O&M Costs
(where CRF = Capital Recovery Factor)

Levelized Cost ($/tonne CO2) = Total Annual Cost ($/yr) / { m * CF * 365 }
(where m = CO2 mass flow rate [tonnes/day], CF = plant capacity factor, and 365 = days per year)

The IEA GHG PH4/6 study reports all cost figures in year 2000 US dollars.

The IEA GHG 2005/2 Models

In 2005, the IEA Greenhouse Gas R&D Programme released two additional, related reports (one for Europe and another for North America) in which the costs and potential of CO2 transport and storage for each of the respective regions were studied. The IEA GHG 2005/2 study focused on Europe. (The study on North America will be discussed later in this report.) Work was carried out by the The Netherlands Geological Survey (TNO-NITG), the geological surveys of Britain (BGS) and Denmark/Greenland (GEUS), and Ecofys.

The equation used for calculating pipeline diameter is:

$$D = [\ m / (0.25\ \pi\ \rho\ v\ )\ ]^{0.5} / 0.0254$$

(where D = pipeline diameter [in], m = CO2 mass flow rate [kg/s], ρ = CO2 density [kg/m³], v = flow velocity [m/s])

In this equation, the study assumes that the flow velocity (v) is a constant 2.0 m/s.

The equation used for calculating onshore pipeline capital costs in the IEA GHG 2005/2 study is taken almost directly from the IEA GHG PH4/6 study for ANSI Class 1500# pipe—the only differences being the following: (1) costs are expressed in euros (€) in the IEA GHG 2005/2 study; (2) a change of sign on the final constant (from +0.000007 to -0.000007), which makes virtually no difference in calculated cost; and (3) an omission of the location factor term, F_L, presumably because in the IEA GHG PH4/6 study $F_L = 1.0$ for Europe, the only region considered in the IEA GHG 2005/2 study. The equation is shown below:

$$\text{Pipeline Capital Cost (€)} = F_T * 10^6 * [\ (0.057 * L + 1.8663) + (0.00129 * L) * D + (0.000486 * L - 0.000007) * D^2\ ]$$

(where F_T = terrain factor, L = pipeline length [km], and D = pipeline diameter [in])

As with the IEA GHG PH4/6 study, terrain factors (F_T) are as follows: cultivated land=1.10, grassland=1.00, wooded=1.05, jungle=1.10, stony desert=1.10, <20% mountainous=1.30, >50% mountainous=1.50. But for the IEA GHG 2005/2 study, an average value of 1.20 is taken for F_T.

For booster stations, capital costs are assumed to be for the most part independent of CO2 mass flow rate, and are instead expressed on a per-kilometer basis. The capital cost equation for onshore booster stations is:

$$\text{Booster Station Capital Cost (€)} = (35{,}000\ \text{€/km}) * L$$

(where L = pipeline length [km])

Hence, the total capital cost is given by:

$$\text{Total Capital Cost (€)} = \text{Pipeline Capital Cost} + \text{Booster Station Capital Cost}$$

It appears that in the IEA GHG 2005/2 report the capital cost of booster stations is always included in the total capital cost, regardless of the presence or absence of a booster station. A short pipeline (e.g., 100 km), however, may not require booster stations. Therefore, when replicating the IEA GHG 2005/2 models, we assume that booster stations are unnecessary if the pipeline length is less than 200 km, which means that the booster station capital cost is not included in the total capital cost. This assumption for a minimum distance of 200 km is consistent with IEA GHG 2005/2 study's own assumption of 200 km for the average distance between two booster stations, which they use for determination of booster station power use.

The equation for booster station pumping power use is:

$$P_p = [(1/\rho) * (\Delta P/\eta_p)] / Dist_{BS}$$

(where P_p = pump power use [J/km/kg CO2], ρ = CO2 density [kg/m^3], ΔP = pressure increase [Pa], η_p = pump efficiency, and $Dist_{BS}$ = average distance between two booster stations [km])

In the IEA GHG 2005/2 study, the following values are assumed for the above equation: $\rho = 800$ kg/m^3, $\Delta P = 4 \times 10^6$ Pa, $\eta_p = 0.75$, and $Dist_{BS} = 200$ km.

The total capital cost is annualized with a 10% discount rate over a 20 year operational lifetime by the following equation:

$$Annual\ Capital\ Cost\ (euros/yr) = Total\ Capital\ Cost\ (euros) \Big/ \frac{(1+i)^n - 1}{i(1+i)^n}$$

(where n = operational lifetime [years], and i = discount rate)

The annual O&M costs of the pipeline are calculated as 3% of the pipeline capital cost. And the annual O&M costs of the booster station are calculated as 5% of the booster station capital cost.

Annual Pipeline O&M Costs (€/yr) = (Pipeline O&M factor) * Pipeline Capital Cost
(where Pipeline O&M factor = 3%)

Annual Booster Station O&M Costs (€/yr) = (Booster Station O&M factor) * Booster Station Capital Cost
(where Booster Station O&M factor = 5%)

The total annual cost is found by summing the annual capital and O&M costs.

Total Annual Cost (€/yr) = Annual Capital Cost
+ Annual Pipeline O&M Costs
+ Annual Booster Station O&M Costs

Finally the levelized cost of CO2 transport is calculated as a combination of the total annual costs and the booster station power required for pumping.

Levelized Cost (€/tonne CO2) = 1000 * { [Total Annual Cost / (m * (31,536,000) * CF)]
+ [COE * P_p * L / (3.6 * 10^6)] }

(where 1000 = kg/tonne, m = CO2 mass flow rate [kg/s], 31,536,000 = seconds per year, CF = plant capacity factor, COE = cost of electricity [€/kWh], P_p = pump power use [J/km/kg CO2], L = pipeline length [km], and 3.6×10^6 = J/kWh)

An electricity cost of 0.04 €/kWh is assumed in the original study.

Finally, the IEA GHG 2005/2 study reports all costs in year 2000 euros.

The IEA GHG 2005/3 Models

As mentioned previously, the IEA GHG 2005/3 study was published in 2005 and focuses on the costs and potential of CO2 transport and storage in North America (onshore USA and Canada). Yet, although the goals of this study were the same as those of the European study (IEA GHG 2005/2), some of the approaches, assumptions, models, and, thus, results differ in marked ways. Work on the North American study was carried out by Battelle and the Alberta Energy and Utilities Board.

To calculate the pipeline diameter, the IEA GHG 2005/3 study cites a rule of thumb in [14] that says the CO2 volumetric flow rate should be 0.65×10^6 scf/day/in^2 of pipe area. In different units, this rule of thumb can be expressed as (18.41 ρ) tonnes/day/in^2 (where ρ is the CO2 density under standard conditions). The pipeline diameter can then be found by:

$$D = \left[\frac{4\ \mathrm{m}}{18.41\ \rho_N\ \pi} \right]^{1/2}$$

(where D = pipeline diameter [in], m = CO2 mass flow rate [tonnes/day], and ρ = CO2 density [kg/Nm^3])

To calculate CO2 pipeline costs, this study takes a similar approach to the MIT study by using historical cost data for natural gas pipeline construction, as reported in the *Oil & Gas Journal*. From this data, they conclude that, on average, construction costs for CO2 pipelines would be $41,681/in/mile ($25,889/in/km). (For comparison, the MIT study concludes that the pipeline cost would $20,989/in/km, as mentioned previously.) In terms of CO2 mass flow rate, the pipeline capital cost is calculated by:

Pipeline Capital Cost ($/mile) = 39,409 * $(m / 24)^{0.5}$
(where m = CO2 mass flow rate [tonnes/day], and 24 = hours/day)

Annualizing the pipeline capital cost with a 10% discount rate over a 25 year operational lifetime yields the following equation:

Annual Pipeline Capital Cost ($/mile/yr) = 4,335 * $(m / 24)^{0.5}$
(where m = CO2 mass flow rate [tonnes/day], and 24 = hours/day)

By assuming the annual O&M costs are 2% of the pipeline capital cost and dividing the total annual capital and O&M costs by the annual CO2 mass flow rate, the total levelized capital and O&M cost equation is given by:

Total Levelized Capital and O&M Cost ($/mile/tonne CO2) = 5123 * $(m / 24)^{0.5}$ / (m * CF * 365)
(where m = CO2 mass flow rate [tonnes/day], 24 = hours/day, CF = plant capacity factor, 365 = days/year)

The levelized costs of CO2 transport is given by:

Levelized Cost ($/tonne CO2) = (L + 10) x 1.17 * (Total Levelized Capital and O&M Cost)
(where L = pipeline length [miles], 10 = extra pipeline distance at injection site [miles],
1.17 = straight line distance adjustment factor)

Finally, since the IEA GHG 2005/3 study does not state the reference year that they express costs in, we assume that they use year 2002 dollars, owing to the fact that 2002 is the year for which they obtained natural gas pipeline cost data from the *Oil & Gas Journal*.

The Parker Models

Like the MIT and IEA GHG 2005/3 studies, Parker uses natural gas pipeline costs, as reported in the *Oil & Gas Journal* for the years 1991-2003. Parker goes further than the other studies, however. Instead of simply reporting one cost (e.g., $/in/km), he fits second order equations to the cost data and develops equations that predict the costs for each of the four different cost categories—materials, labor, miscellaneous, and right of way. The equations are functions of pipeline diameter and length. And although Parker's pipeline cost equations were developed with the intent of predicting costs for hydrogen pipelines, they can still be used for CO2 pipelines, as is done in the MIT and IEA GHG 2005/3 studies.

The pipeline capital cost equations of Parker are shown below:

$$\text{Materials Cost (\$)} = [330.5 * D^2 + 687 * D + 26{,}960] * L + 35{,}000$$

$$\text{Labor Cost (\$)} = [343 * D^2 + 2{,}074 * D + 170{,}013] * L + 185{,}000$$

$$\text{Miscellaneous Cost (\$)} = [8{,}417 * D + 7{,}324] * L + 95{,}000$$

$$\text{Right of Way Cost (\$)} = [577 * D + 29{,}788] * L + 40{,}000$$

$$\begin{aligned}\text{Total Capital Cost (\$)} &= \text{Materials Cost} + \text{Labor Cost} + \text{Miscellaneous Cost} + \text{Right of Way Cost}\\ &= [673.5 * D^2 + 11{,}755 * D + 234{,}085] * L + 355{,}000\end{aligned}$$

(where D = pipeline diameter [in], and L = pipeline length [miles])

Two of the above equations have been slightly adapted. First, consider the equation for 'Right of Way Cost'. On page 17 of his report, Parker states that the diameter term in '577 * D' should be squared, '577 * D^2', however, we believe this to be a typographical error, since the regression equation on Figure 18 of the same page shows the term to be '576.78 * D' (i.e., without the squared exponent). Furthermore, the '577 * D' term (unsquared) is evidently the one that is used when adding up the four individual equations to generate the 'Total Capital Cost' equation. Correcting for this error has large implications on the calculated right of way cost, potentially increasing it by a factor of ten for large diameter pipelines. Second, consider the 'Total Capital Cost' equation. Our equation is nearly identical to that of Parker, aside from the final term and some small rounding differences on the first and second terms. We have added up the final terms from each of the four individual equations (35,000 + 185,000 + 95,000 + 40,000) to get 355,000, which we use in our equation for total capital cost, as compared to the 405,000 term that Parker uses. These latter differences are, of course, much smaller in importance than those in the equation for right of way costs. Recent discussions with Parker have confirmed the presence of the typographical errors. Our adaptation of his methodology is, therefore, justified.

Parker reports all costs in year 2000 dollars.

SUMMARY AND COMPARISON OF MODELS

The above descriptions of the various CO2 transport models show that each study takes a somewhat unique approach in sizing the pipeline and estimating the associated costs. Further, each of the various studies is built upon differing assumptions and input, which in turn leads to dissimilar output. The main differences in the studies are highlighted in the following table. (Note that the Parker study is unique in that it does not deal strictly with CO2 transport and sequestration and, thus, cannot be compared to the other studies on many accounts.)

	Study						
	Ogden	MIT	Ecofys	IEA GHG PH4/6	IEA GHG 2005/2	IEA GHG 2005/3	Parker
Reference Cost Year	2001	1998	2003	2000	2000	2002	2000
Capital Recovery Factor [%/yr]	15	15	--	'user specified'	--	--	--
Discount Rate [%]	--	--	10	--	10	10	--
Operational Lifetime [years]	--	--	25	--	20	25	--
O&M Factor	4.0%/year of total capital cost	$3,100/km/year	2.1%/year of total capital cost	'by equation'	3%/yr of pipeline capital + 5%/yr of booster capital	2.0%/year of total capital cost	--
Plant Capacity Factor [%]	--	80	'not reported'	'user specified'	90	'not reported'	--
Electricity Cost [/kWh]	--	--	--	'user specified'	0.04 €	--	--
Booster Stations Included?	No	No	No	Yes / No (user specified)	Yes	No	--
Pipeline Inlet Pressure [MPa]	15	15.2	12	'user specified'	--	--	--
Pipeline Outlet Pressure [MPa]	10	10.3	8	'user specified'	--	--	--
Friction Factor, f	~0.0021 (by equation)	~0.0033 (by Moody chart)	0.015 - 0.020 (= 4 x f)	0.015 (= 4 x f)	--	--	--
CO2 Temperature [^{o}C]	4.44 - 37.78	25	10	--	--	--	--
CO2 Density [kg/m^3]	--	884	800	800	800	--	--
CO2 Viscosity [N-s/m^2]	--	6.06 x 10^{-5}	--	--	--	--	--
Terrain Factor	--	--	1.0	1.05 - 1.50 (by terrain)	1.05 - 1.50 (by terrain)	1.17	--
Location Factor	--	--	--	0.7 - 1.2 (by location)	--	--	--

Table 1: Assumptions and input for each of the various studies

For starters, the various studies report their costs based in different reference years, and as shown previously, two of the studies (Ecofys and IEA GHG 2005/2) report costs in euros, not dollars. In addition, some studies annualize their capital costs by a simple capital recovery factor (CRF), while others annualize by discount rate and project lifetime. For comparison, annualizing with a 10% discount rate over 20 years is similar to using a CRF of 11.7%; annualizing with a 10% discount rate over 25 years is similar to using a CRF of 11.0%. In other words, the studies that annualize capital costs by discount rate and lifetime inherently estimate annual capital costs that are lower than those estimated by studies that utilize the CRF method. On a similar subject, there is variation between the studies in the annual O&M cost factor that is applied. This variation is by no means trivial, considering that in the range of CO2 mass flow rates and pipeline sizes and lengths that would generally be required for CO2 sequestration, the O&M factors for both the MIT and IEA GHG PH4/6 studies would be in the range of 0.60% to 3%.

As seen in the table, the IEA GHG PH4/6 and IEA GHG 2005/2 studies are the only ones that include booster stations in their estimation of capital costs. The requirement of booster stations could potentially have a large effect costs, possibly leading these studies to report CO2 transport costs that are much higher than other studies. Other variables that may lead to differing cost estimates are the assumed fluid flow properties of CO2—inlet and outlet pressure, friction factor, density—and the method for calculating pipeline diameter, which is based on these flow properties. While all studies recognize that CO2 pressure should stay above the critical pressure of 7.38 MPa, they differ in their assumption of how low it can drop. This assumption is likely based on whether or not the CO2 is assumed to be recompressed at the injection site or if it will be directly injected via the outlet pipeline pressure. Notice, however, that two of the studies (IEA GHG 2005/2 and IEA GHG 2005/3) do not even consider inlet and outlet pressure when sizing/costing the pipeline. But of the ones that do consider pressure, the assumed pressure drop is between 4 and 5 MPa. Moreover, the studies assume different values for friction factor. As seen in the table, the Ecofys and IEA GHG PH4/6 studies use friction factors that are a magnitude of four times greater than the Ogden and MIT studies; though, this difference is canceled out due to the particular equation that is used. Yet, even after accounting for this difference, the friction factors that are calculated and used in the Ogden study are consistently lower than those of other studies. A smaller friction factor will, thus, lead to the calculation of a smaller required pipeline diameter.

Lastly, while only the IEA GHG PH4/6 study uses a location factor when calculating capital costs, four of the studies use a terrain factor. The IEA GHG 2005/2 study simply cites the terrain factors of the IEA GHG PH4/6 study and assumes that a value of 1.20 would be satisfactory. The IEA GHG 2005/3 study on the other hand, proposes an average terrain factor of 1.17 for the varied geography of North America and notes that it is approximately the average of the IEA GHG PH4/6 terrain factors.

APPLYING COMMON BASES TO ALL MODELS

Since each of the studies is based on differing assumptions and input, it is difficult to compare their output on an 'apples to apples' basis. Therefore, we have replicated each of the aforementioned models and put all of them on the same common bases. These bases, which are enumerated in the following table, were chosen as a result of their use in one or more of the original studies.

Common Design Bases	
Plant Capacity Factor [%]	80
Pipeline Inlet Pressure [MPa]	15.2
Pipeline Outlet Pressure [MPa]	10.3
CO2 Temperature [^oC]	25
CO2 density [kg/m^3]	884
CO2 Density @ STP [kg/Nm^3]	1.965
CO2 viscosity [$N\text{-}s/m^2$]	6.06E-05

Common Economic Bases	
Reference Cost Year	2005
Conversion of Euro to Dollar [$/€]	1.20
Operational Lifetime [years]	20
Discount Rate [%]	10
Location Factor	1.00
Terrain Factor	1.20
Electricity Cost [$/kWh]	0.04

Table 2: Common bases used when replicating the various models

For these common bases, we used the models from each of the studies to calculate the pipeline diameter [in], capital cost [$/km], and levelized cost [$/tonne CO2] as a function of both CO2 mass flow rate [tonnes CO2/day] and pipeline length [km]. The range of interest for CO2 mass flow rate is 1,000 to 20,000 tonnes/day, which roughly corresponds to conventional pulverized coal coal plants with a generating capacity between 50 and 1000 MW. For pipeline length, the range is 100 to 500 km. The following graphs show the diameters, capital costs, and levelized costs that are calculated by the various models for a 100 km pipeline at different flow rates. (Exact values can be found in the appendix.) The averages of all the models for the calculated values at each flow rate are also shown on the graphs.

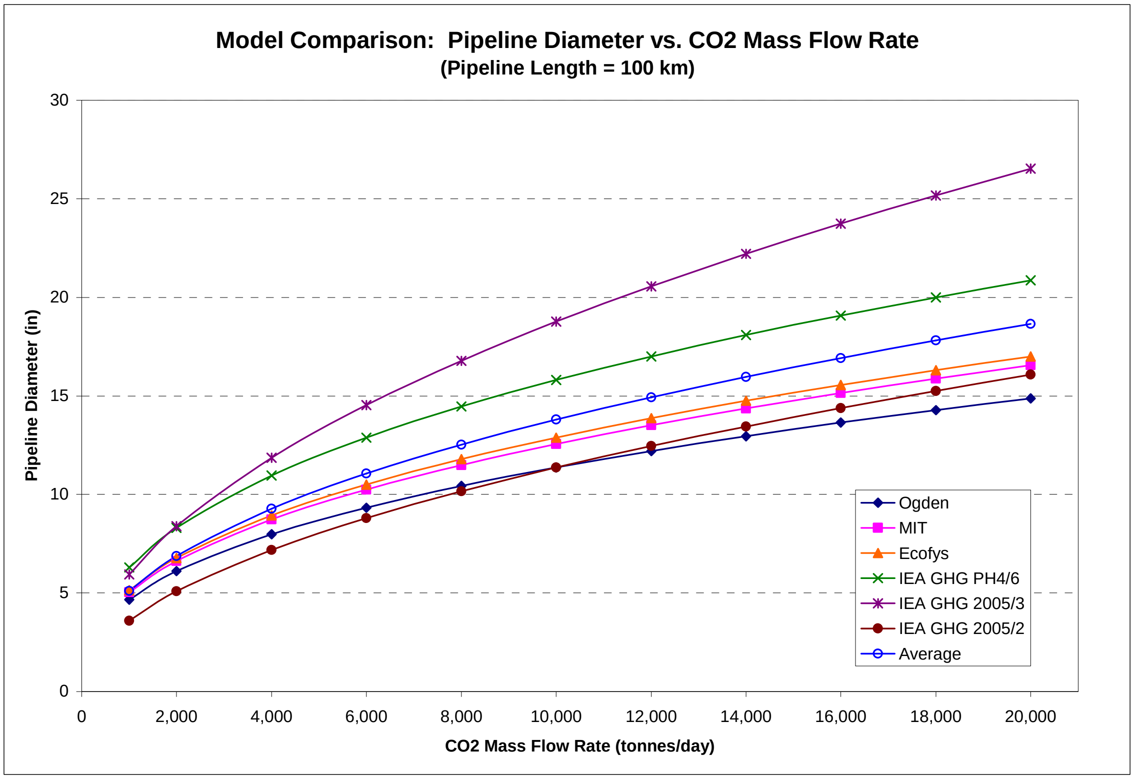


Figure 1: Pipeline Diameter vs. CO2 Mass Flow Rate (Pipeline Length = 100 km)

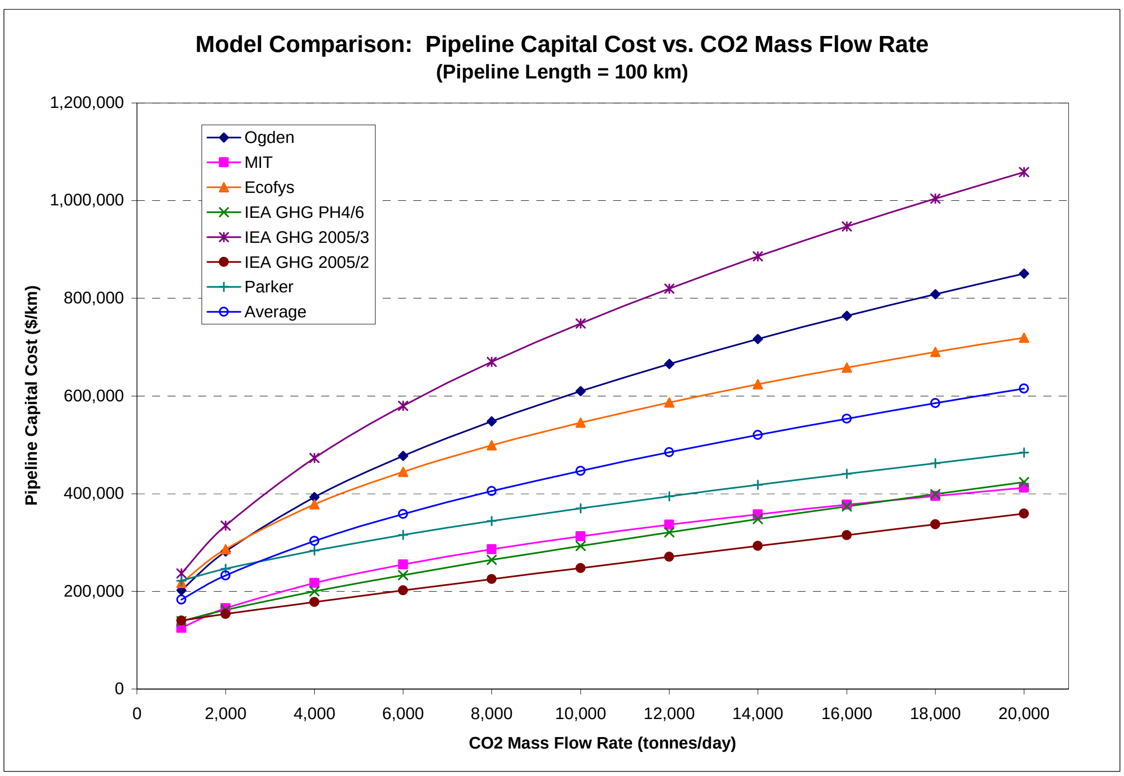


Figure 2: Pipeline Capital Cost vs. CO2 Mass Flow Rate (Pipeline Length = 100 km)

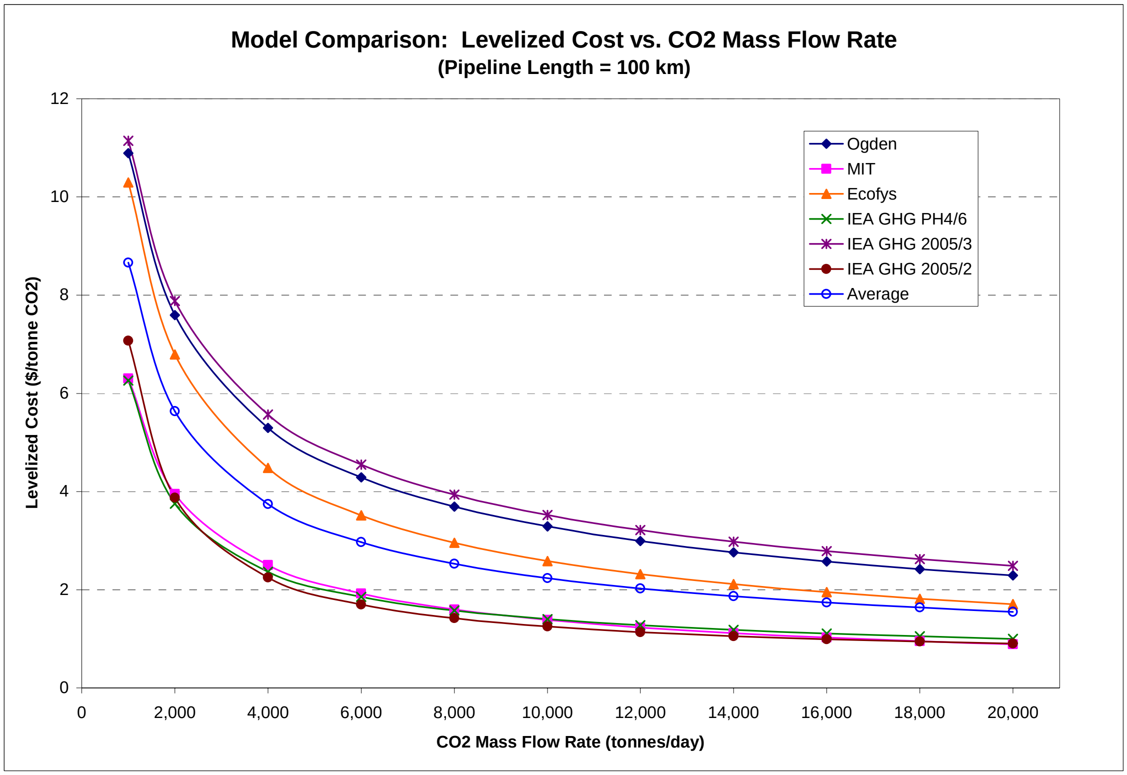


Figure 3: Levelized Cost vs. CO2 Mass Flow Rate (Pipeline Length = 100 km)

The graphs show that each of the models exhibits the same trends: as CO2 mass flow rate gets larger, pipeline diameter and capital cost increase, whereas levelized cost decreases. In addition, at higher flow rates the differences between the models increase for diameter and capital cost, but decrease for levelized cost. (When calculating capital cost with the Parker models, which have no equations for calculating pipeline diameter, we use a diameter at each of the different mass flow rates that is an average of the other models.)

Comparing each of the models' calculated pipeline diameters to the average, it is evident that IEA GHG 2005/3 and IEA GHG PH4/6 models consistently estimate diameters that are above the average, while the Ecofys, MIT, Ogden, and IEA GHG 2005/2 models consistently estimate diameters that are below average. One would expect that a below average estimated pipeline diameter would lead to a below average estimated capital cost, but this is not necessarily the case. For example, the Ecofys and Ogden models estimate diameters that are below average but estimate capital costs that are above average. Conversely, although the IEA GHG PH4/6 estimates an above average pipeline diameter, it estimates capital cost below the average. Finally, the trends on the levelized cost graph are almost the same as those on the capital cost graph, with the IEA GHG 2005/3, Ogden, and Ecofys models estimating levelized costs that are above average, and the MIT, IEA GHG PH4/6, and IEA GHG PH 2005/2 estimating costs that are below average.

One question that arises from these comparisons relates to the effect of assumed friction factor on the calculated pipeline diameter of each of the models. In the above graphs, we let each model retain its own original friction factor. As discussed previously, the Ogden and MIT models both use equations for estimating a reliable friction factor; for Ogden the value is generally about 0.0021, and for MIT it is about 0.003-0.004 in the range of flow rates considered here. The Ecofys and IEA GHG PH4/6 models assume constant friction factors that are 0.015/0.020 and 0.015, respectively; but remember that their values are a magnitude of four times greater than those of Ogden and MIT due to the type of fluid flow equation that they use. (Note that we chose an Ecofys friction factor of 0.015, instead of 0.020, when calculating diameter for the above graphs.) The IEA GHG 2005/2 and IEA GHG 2005/3 models, on the other hand, do not employ fluid flow equations that are functions of friction factor, so their models are not dependent on the choice of friction factor. Hence, it seems reasonable to assume a friction factor of 0.00375 for Ogden and MIT (i.e., 0.015 for Ecofys and IEA GHG PH4/6) and apply this factor as a common basis across all of the models. Doing this has the effect of increasing the calculated pipeline diameters of both the Ogden and MIT models, while contributing to no change in any of the other models since either the friction factor was unchanged or it is not a part of the fluid flow equations. The following graph shows calculated pipeline diameters for each of the models when a common friction factor is applied; it should be compared to the graph in Figure 1, which shows calculated pipeline diameters for each of the models when their original friction factors are assumed.

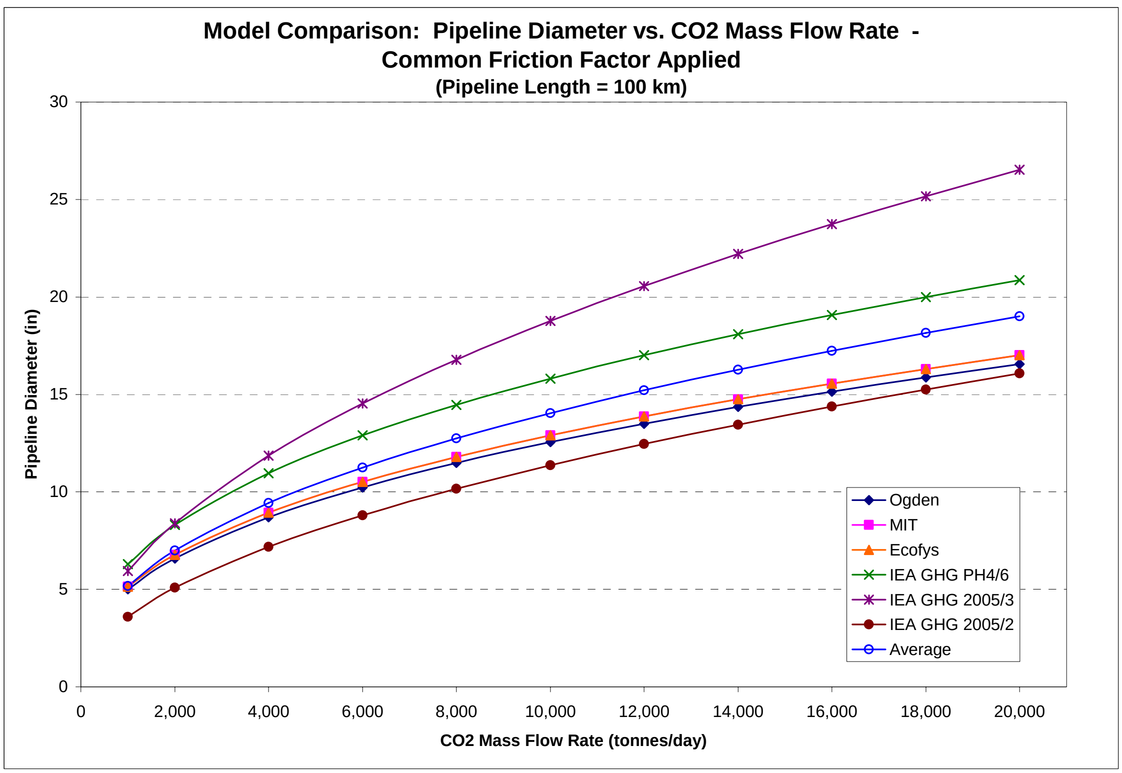


Figure 4: Pipeline Diameter vs. CO2 Mass Flow Rate – Common Friction Factor Applied (Pipeline Length = 100 km)

Since Ogden's friction factors were initially the lowest, its calculated diameters experience the largest change, increasing from 7.1-11.4% over the flow rate range of 1,000 to 20,000 tonnes/day. The MIT diameters also experience a change, albeit to a smaller degree, increasing from 1.8-2.7% over the same flow rate range. Interestingly, since the MIT and Ecofys equations for calculating pipeline diameter are essentially rearranged versions of each other, applying a common friction factor to both of them causes their calculated diameters to be exactly the same.

Further, the average calculated pipeline diameter of all of the models increases by 1.4-1.9% over the flow rate range.

A second question that arises from the comparisons relates to the effect of pipeline diameter on the capital and levelized costs of each of the models. Since each model estimates a different diameter for the same operating conditions, it would stand to reason that this would significantly affect each model's cost estimates. We can test this hypothesis by taking an average of the models' calculated pipeline diameters at each of the various flow rates (assuming that each model uses its own original friction factor) and then applying this average diameter to the capital and level cost equations in all models. This will allow the cost models to be compared on a standard basis. The following two graphs show the capital and levelized costs of each model when a common diameter is applied at each flow rate; it should be compared to the graphs in Figures 2 and 3, which show capital and levelized costs for each of the models when using their originally calculated pipeline diameters. (Exact values can be found in the appendix.)

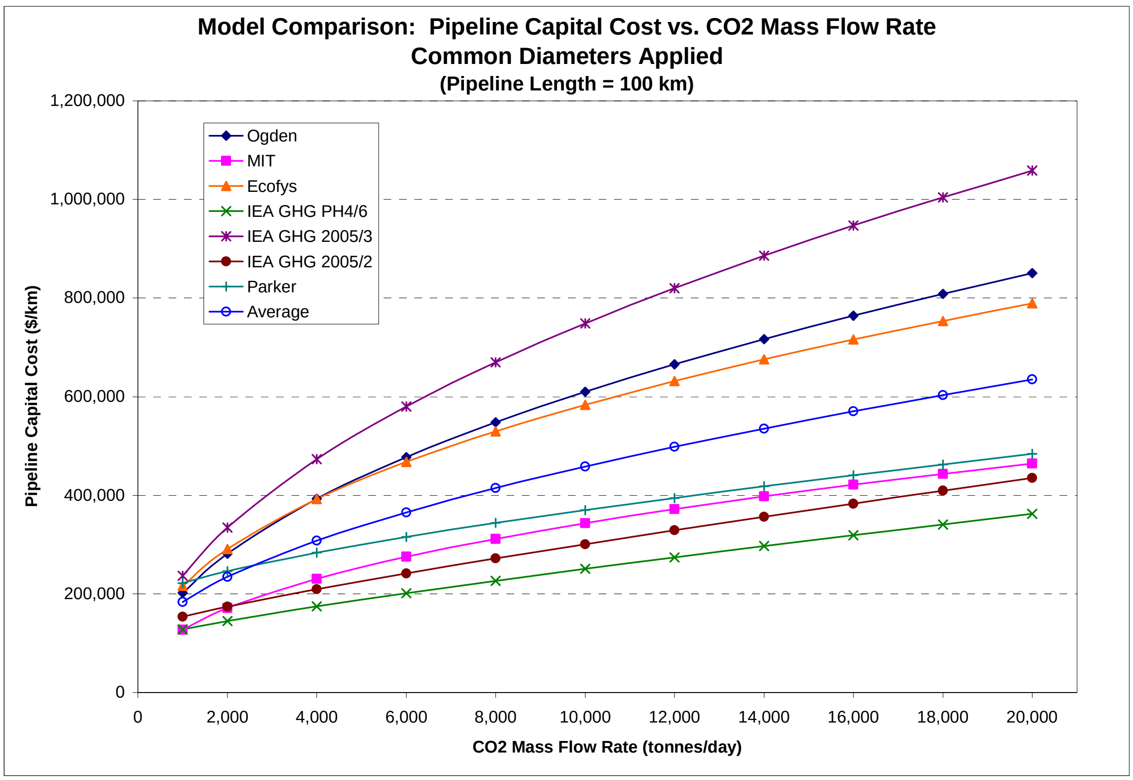


Figure 5: Pipeline Capital Cost vs. CO2 Mass Flow Rate – Common Diameters Applied (Pipeline Length = 100 km)

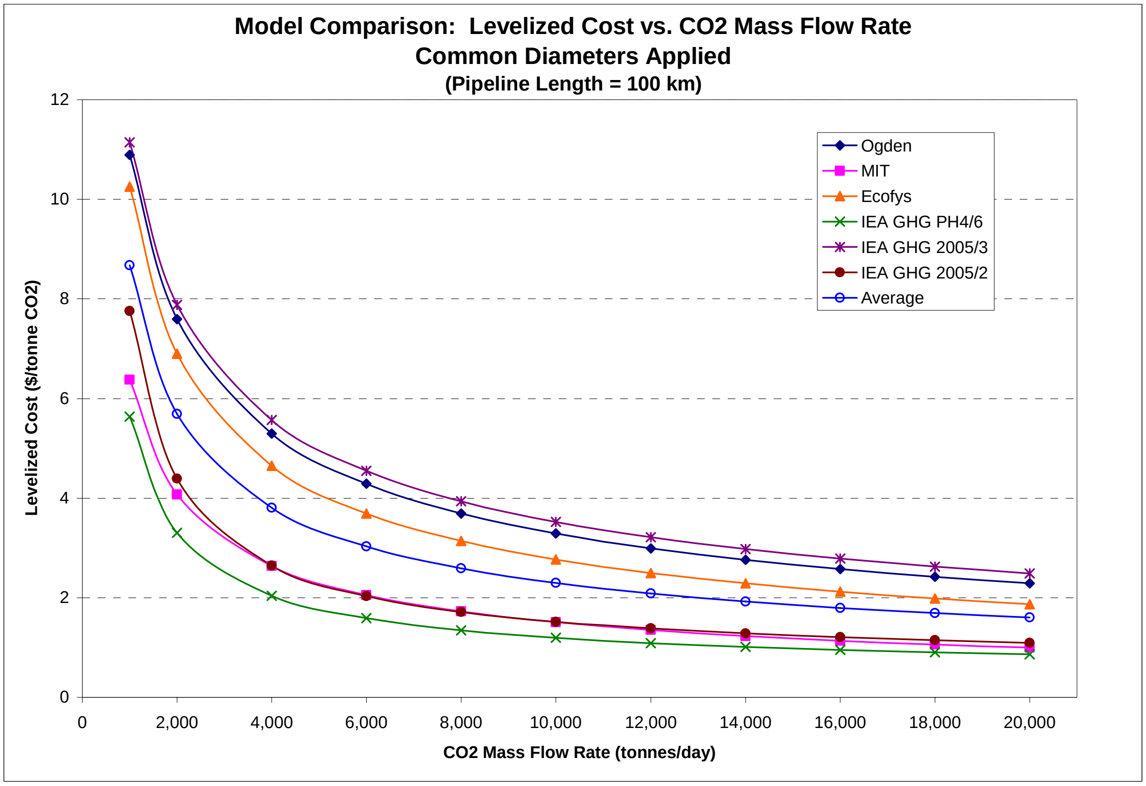


Figure 6: Levelized Cost vs. CO2 Mass Flow Rate – Common Diameters Applied (Pipeline Length = 100 km)

As shown in these two graphs, even when the same pipeline diameters are applied across all models, the variation in cost estimates is almost as high as when different diameters are used. This shows that the variability in the models' cost estimates is not just a function of the diameter they use, but is really more a function of each models' methods and assumptions for calculating costs. For starters, the capital and levelized costs calculated by the IEA GHG 2005/3 models do not change when different diameters are used; this is because their cost equations are not functions of diameter. Similarly, Ogden's costs do not change because her models are not directly functions of diameter. Parker's capital costs also do not change, but this is simply because average diameter was already being used in the cost calculations for his model. The capital and levelized costs of the Ecofys, MIT, and IEA GHG 2005/2 models all increase when a common diameter is applied; this is due to the common diameter at each flow rate being greater than the originally calculated diameters of these models. Conversely, the costs of the IEA GHG PH4/6 models decrease because the common diameters are less than the originally calculated diameters of this model.

The next series of graphs shows for each of the different models the dependence of pipeline diameter, capital cost, and levelized cost on pipeline length at a constant CO2 mass flow rate of 10,000 tonnes/day. This particular flow rate is shown because it is in the middle of the range of interest and because the trends of the models at this flow rate are representative of those at other flow rates.

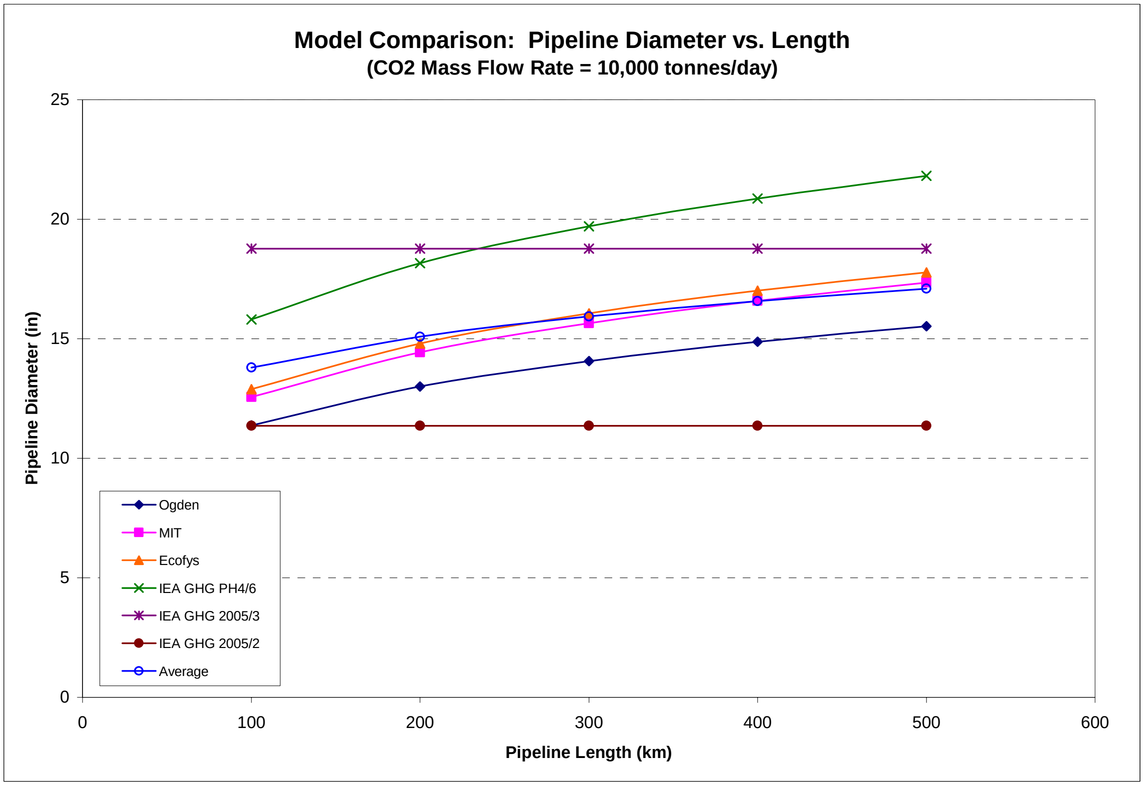


Figure 7: Pipeline Diameter vs. Length (CO2 Mass Flow Rate = 10,000 tonnes/day)

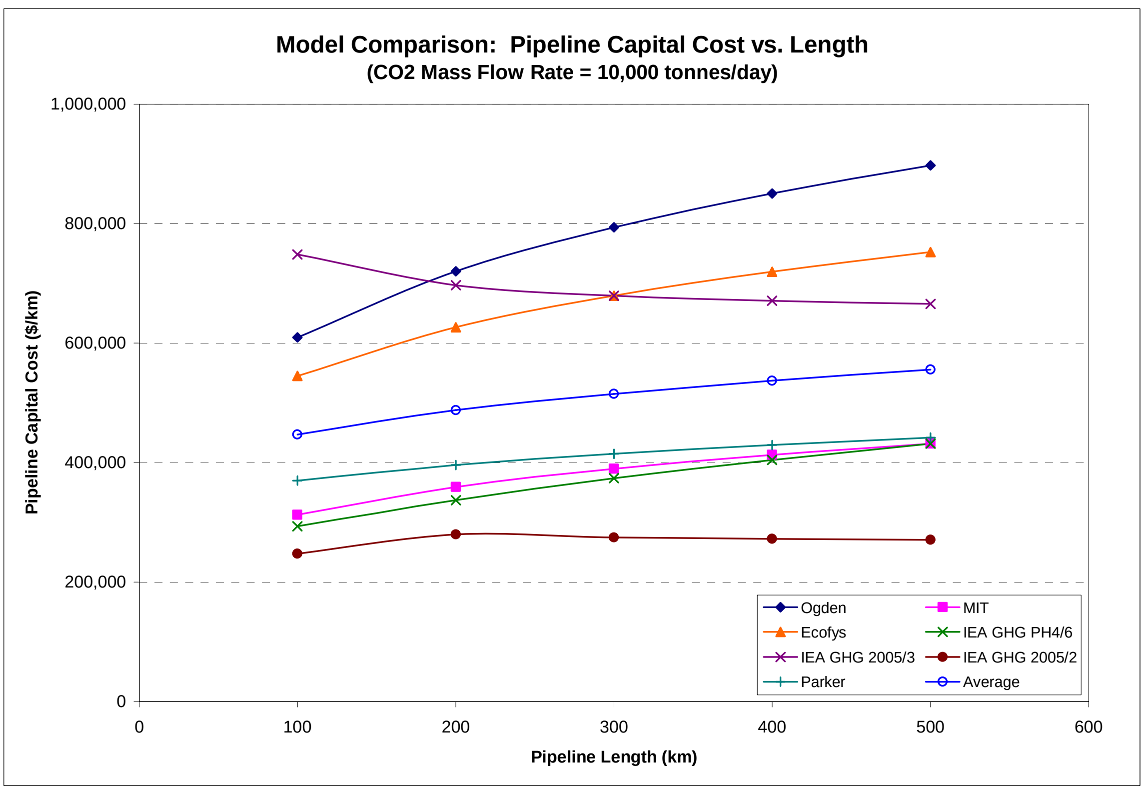


Figure 8: Pipeline Capital Cost vs. Length (CO2 Mass Flow Rate = 10,000 tonnes/day)

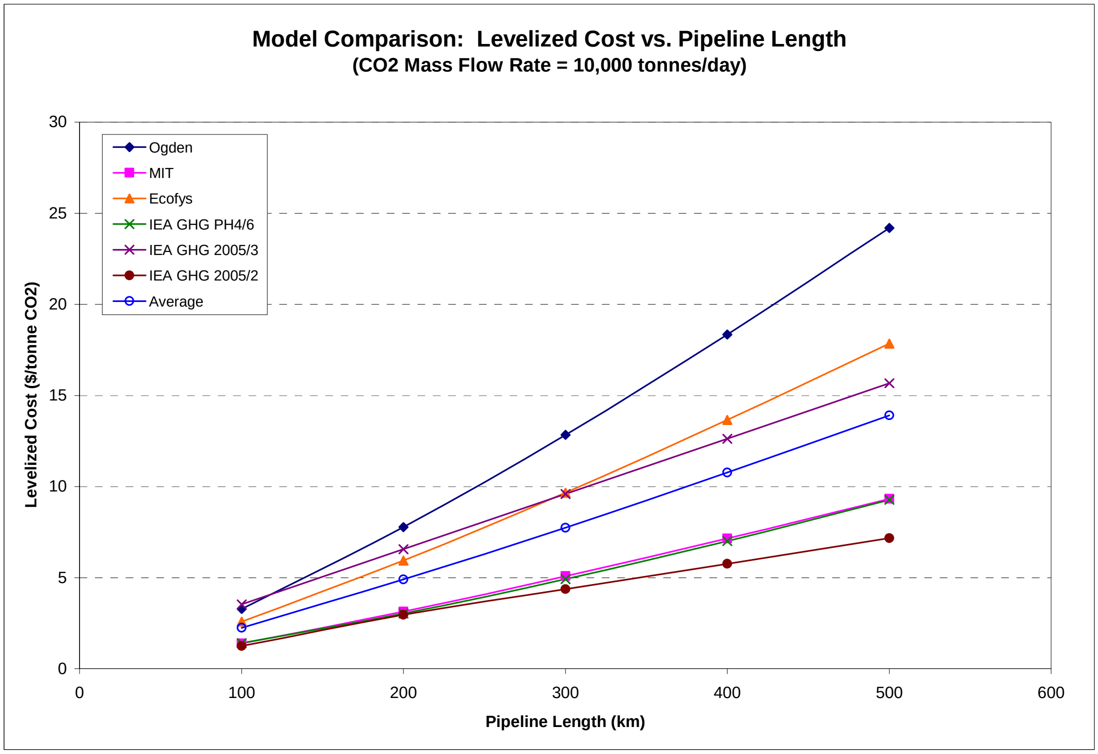


Figure 9: Levelized Cost vs. Pipeline Length (CO2 Mass Flow Rate = 10,000 tonnes/day)

The latter three graphs for dependence of pipeline diameter, capital cost, and levelized cost on pipeline length are unlike the former three graphs for dependence on CO2 mass flow rate in that the trends of the different models are not always the same. For instance, consider pipeline diameter versus length. The diameter generally increases as the pipeline gets longer, except in the IEA GHG 2005/2 and IEA GHG 2005/3 models, where the diameter is constant for all lengths since the equations for calculating diameter in these models are not functions of length. The trends for pipeline diameter also affect the trends for capital cost per unit length. Again, except for the IEA GHG 2005/2 and IEA GHG 2005/3 models, capital cost generally increases as the pipeline gets longer. For the IEA GHG 2005/3 models, the equation for calculating capital cost is not actually a function of diameter, so one would expect that the capital cost would be constant for all pipeline lengths. This is not the case, however, because the pipeline adjustment distance of 10 miles becomes gradually less significant as the pipeline length increases. For the IEA GHG 2005/2 models, the capital cost increases up to a pipeline length of 200 km because, as mentioned previously, when replicating these models we assume that booster stations are not required unless the pipeline length is 200 km or greater. Thus, the inclusion of booster stations at 200 km causes the capital cost to increase, but as the pipeline gets longer, the booster station cost becomes gradually less significant, contributing to the declining capital cost versus pipeline length that is seen in the graph above. Finally, the trends for levelized cost as a function of pipeline length are similar those for levelized cost as a function of CO2 mass flow rate, with the IEA GHG 2005/3, Ogden, and Ecofys models estimating levelized costs that are above average, and the MIT, IEA GHG PH4/6, and IEA GHG PH 2005/2 estimating costs that are below average. Of course, levelized cost dependence on pipeline length is different in that it increases as the pipeline gets longer.

COMBINING THE MODELS

One of the main goals of the current study is to create new models similar to, and based upon, all of the other models that have been described and compared up to this point. This is best achieved by taking averages of the output from the various models at different CO2 mass flow rates and pipeline lengths after they have been put on common bases. Ideally, we would like to create models that, like the Ogden models, are functions only of CO2 mass flow rate and pipeline length, not pipeline diameter. The graphs above show the model averages for pipeline diameter, capital cost, and levelized cost for a few flow rates and pipeline lengths, but calculations at many more operating points have also been carried out. The results for capital cost are shown in the following graph.

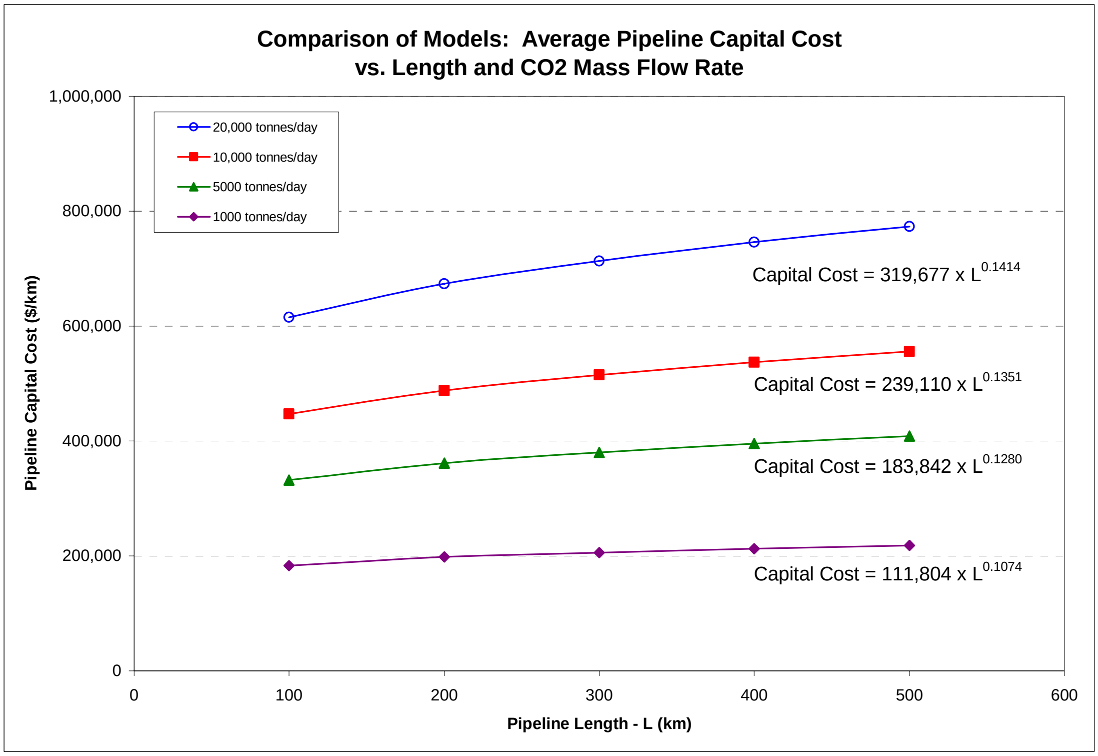


Figure 10: Average Pipeline Capital Cost vs. Length and CO2 Mass Flow Rate

The above graph clearly shows that the capital cost per unit length increases with both pipeline length and CO2 mass flow rate. In addition, the dependence on length appears to increase at higher flow rates. This is confirmed by comparing the best-fit power regression equations at each of the different flow rates. These regression equations can be generically written as:

Pipeline Capital Cost [$/km] = (constant) * $L^{exponent}$
(where L = pipeline length [km])

The exponent on the length term (L) increases as flow rate increases, but it is approximately 0.13 in the range of flow rates that will typically be encountered in CCS applications. Similarly, the constant that precedes the length term also increases as flow rate increases. This dependence is best described by the following exponential equation:

constant = 9970 * $m^{0.35}$
(where m = CO2 mass flow rate [tonnes/day])

Therefore, by substituting for the exponent and constant terms, the generic equation for pipeline capital cost can be rewritten as the following:

Pipeline Capital Cost [$/km] = (constant) * $L^{exponent}$
= (9970 * $m^{0.35}$) * $L^{0.13}$
(where m = CO2 mass flow rate [tonnes/day], and L = pipeline length [km])

The given costs are in year 2005 US dollars. This equation provides a way of estimating the pipeline capital cost per unit length based on two quantities that are typically known—CO2 mass flow rate and pipeline length. By approaching the capital cost in this way, one can avoid the need for calculating the pipeline diameter in advance. As previously shown, different models use different methods for calculating pipeline diameter, and more often than not, the end results vary. Moreover, diameters that are significantly different can potentially lead to capital and levelized costs that also differ markedly. For these reasons, it becomes advantageous to estimate capital costs from quantities other than diameter. The above equation is a reliable method for calculating pipeline capital cost since it is essentially derived from seven other pipeline models, all of which are recent and reliable. It should be noted, however, that the equation is based on a set of common bases and assumptions that may be different in practice. Yet, this set of bases is also derived from the other models, and the bases that have ultimately been chosen are ones that are usually used in the context of studying CO2 transport.

As seen in the earlier graphs in this report, however, there is considerable variation for the pipeline capital costs that are calculated by the models of the different studies. Therefore, to put this study's pipeline capital cost equation into context, it is helpful to show the range of high and low values from other studies.

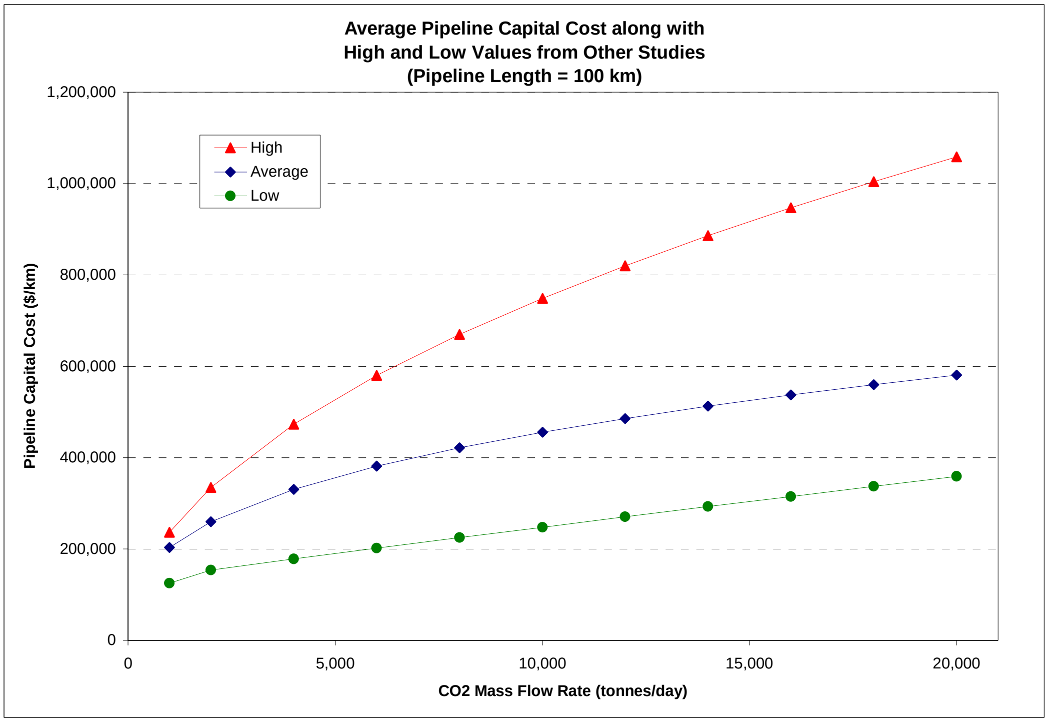


Figure 11: Average Pipeline Capital Cost Compared to High and Low Estimates from Other Studies

The low cost value at a given flow rate is simply the minimum value that is calculated by any one of the studies. The high values are found in a similar manner by finding the maximum. It is interesting to note that while one study may calculate the maximum/minimum cost at one flow rate, another study may calculate the maximum/minimum at another flow rate. The equations for the low and high values are given roughly as:

$$\text{Pipeline Capital Cost (Low) [\$/km]} = (\text{constant}) * L^{\text{exponent}}$$
$$= (8500 * m^{0.35}) * L^{0.06}$$

$$\text{Pipeline Capital Cost (High) [\$/km]} = (\text{constant}) * L^{\text{exponent}}$$
$$= (4100 * m^{0.50}) * L^{0.13}$$

(where m = CO2 mass flow rate [tonnes/day], and L = pipeline length [km])

Notice that the high and low equations are not simple multiples of the average cost equation, just like the high, low, and average cost lines in the above graph are not 'parallel' to each other. In this sense, the high and low equations/lines should not be considered similar to a confidence interval, equally bounding the upper and lower ranges of the average. Rather, the high and low simply represent the maximum/minimum cost at a given flow rate that was calculated by any of the studies that were considered in this report. Therefore, one can feel comfortable in knowing that while the uncertainty of CO_2 pipeline capital cost may be great, it will very likely be within these high and low bounds.

CONCLUSION

This report has shown that the uncertainty in calculating the costs of CO_2 pipeline transport are quite high, despite the fact that significant field experience already exists in the oil and gas industry. The wide variability in cost estimates results from the different approaches and models that the various studies use, as well as the different input that they assume. In this report, we have attempted to cancel out the effects of differing input assumptions by putting all of the models on the same common bases and comparing them "apples-to-apples". But even after doing this, variability is still high.

Moreover, we have provided a new equation for calculating pipeline capital cost that is based on a handful of other models that have come out in recent years. We have also given equations for the upper and lower bounds that CO_2 pipeline capital costs should be expected to fall within; again, these equations are also based on other recent models. Ultimately, the best way to reduce the variability in cost estimates is through more real-world experience with carbon dioxide capture, transport, and storage—something that we should be able to do based on the few CCS pilot projects that will be carried out across the world in the near future.

REFERENCES

[1] Ogden, J., C. Yang, N. Johnson, J. Ni, J. Johnson, “Conceptual Design of Optimized Fossil Energy Systems with Capture and Sequestration of Carbon Dioxide,” Report to the U.S. Department of Energy National Energy Technology Laboratory (September 30, 2004).

[2] Heddle, G., H. Herzog, M. Klett, “The Economics of CO2 Storage,” MIT LFEE 2003-003 RP (August 2003).

[3] Hendriks, N., T. Wildenborg, P. Feron, W. Graus, R. Brandsma, “EC-Case Carbon Dioxide Sequestration,” M70066, Ecofys (December 2003).

[4] IEA Greenhouse Gas R&D Programme, “Transmission of CO2 and Energy,” Report no. PH4/6 (March 2002).

[5] IEA Greenhouse Gas R&D Programme, “Building the Cost Curves for CO2 Storage: European Sector,” Report no. 2005/2 (February 2005).

[6] IEA Greenhouse Gas R&D Programme, “Building the Cost Curves for CO2 Storage: North America,” Report no. 2005/3 (February 2005).

[7] Parker, N., “Using Natural Gas Transmission Pipeline Costs to Estimate Hydrogen Pipeline Costs,” UCD-ITS-RR-04-35, Institute of Transportation Studies – University of California, Davis, (2004)

[8] Farris, C.B., “Unusual Design Factors for Supercritical CO2 Pipelines,” *Energy Progress*, 3, 150-158 (1983).

[9] Skovholt, O., "CO2 Transportation System," *Energy Conservation and Management*, 34 (9-11): 1095-1103 (1993).

[10] Herzog, Howard, personal communication (January 30, 2006).

[11] Fox, Chuck, e-mail communication to MIT (December 30, 1999).

[12] Crane, “Flow of Fluids through valves, fittings, and pipe,” Technical Paper No. 410M (1982).

[13] Gas Processors Suppliers Association “Engineering Data Handbook,” 10^{th} Ed. (1987).

[14] Brown, D.R., K.K. Humphreys, and LW Vail, “Carbon Dioxide Control Cost for Gasification Combined-Cycle Power Plants in the United States,” PNL-SA-22634, Pacific Northwest Laboratory, Richland, Washington (1993).

APPENDIX

Pipeline Diameter, Capital Cost, and Levelized Cost as a Function of CO2 Mass Flow Rate (Pipeline Length = 100 km)

CO2 Mass Flow Rate (tonnes/day)	Key Output	Model						
		Ogden	*MIT*	*Ecofys*	*IEA GHG PH4/6*	*IEA GHG 2005/3*	*IEA GHG 2005/2*	*Parker*
1,000	Pipeline Diameter (in)	4.67	5.04	5.13	6.29	5.93	3.59	5.11
	Pipeline Capital Cost ($/km)	201,941	125,437	217,078	138,925	236,698	140,081	221,547
	Levelized Cost ($/tonne CO2)	10.89	6.31	10.29	6.26	11.14	7.07	--
2,000	Pipeline Diameter (in)	6.10	6.63	6.77	8.31	8.39	5.08	6.88
	Pipeline Capital Cost ($/km)	281,656	164,990	286,436	161,661	334,741	153,346	246,043
	Levelized Cost ($/tonne CO2)	7.59	3.95	6.79	3.75	7.88	3.87	--
4,000	Pipeline Diameter (in)	7.98	8.72	8.93	10.96	11.87	7.19	9.27
	Pipeline Capital Cost ($/km)	392,838	217,203	377,955	199,773	473,396	178,056	283,792
	Levelized Cost ($/tonne CO2)	5.30	2.50	4.48	2.36	5.57	2.25	--
6,000	Pipeline Diameter (in)	9.33	10.25	10.51	12.89	14.53	8.80	11.05
	Pipeline Capital Cost ($/km)	477,241	255,183	444,505	233,282	579,789	201,743	315,395
	Levelized Cost ($/tonne CO2)	4.29	1.92	3.51	1.85	4.55	1.70	--
8,000	Pipeline Diameter (in)	10.43	11.49	11.79	14.46	16.78	10.17	12.52
	Pipeline Capital Cost ($/km)	547,909	286,126	498,714	264,183	669,483	224,901	343,740
	Levelized Cost ($/tonne CO2)	3.69	1.60	2.96	1.58	3.94	1.42	--
10,000	Pipeline Diameter (in)	11.37	12.56	12.89	15.81	18.76	11.37	13.79
	Pipeline Capital Cost ($/km)	609,853	312,707	545,275	293,310	748,505	247,721	369,864
	Levelized Cost ($/tonne CO2)	3.29	1.38	2.59	1.40	3.52	1.25	--
12,000	Pipeline Diameter (in)	12.20	13.51	13.86	17.01	20.55	12.45	14.93
	Pipeline Capital Cost ($/km)	665,629	336,259	586,527	321,110	819,946	270,301	394,605
	Levelized Cost ($/tonne CO2)	2.99	1.23	2.32	1.28	3.22	1.14	--
14,000	Pipeline Diameter (in)	12.95	14.36	14.75	18.09	22.20	13.45	15.97
	Pipeline Capital Cost ($/km)	716,748	357,558	623,831	347,866	885,643	292,698	418,241
	Levelized Cost ($/tonne CO2)	2.76	1.12	2.11	1.18	2.98	1.06	--
16,000	Pipeline Diameter (in)	13.64	15.15	15.55	19.08	23.73	14.38	16.92
	Pipeline Capital Cost ($/km)	764,192	377,102	658,057	373,762	946,792	314,952	440,721
	Levelized Cost ($/tonne CO2)	2.58	1.03	1.95	1.11	2.79	0.99	--
18,000	Pipeline Diameter (in)	14.28	15.87	16.30	20.00	25.17	15.25	17.81
	Pipeline Capital Cost ($/km)	808,641	395,230	689,803	398,933	1,004,224	337,087	462,551
	Levelized Cost ($/tonne CO2)	2.42	0.95	1.82	1.05	2.63	0.95	--
20,000	Pipeline Diameter (in)	14.87	16.56	17.01	20.86	26.53	16.07	18.65
	Pipeline Capital Cost ($/km)	850,588	412,186	719,495	423,483	1,058,545	359,124	483,837
	Levelized Cost ($/tonne CO2)	2.29	0.89	1.71	1.00	2.49	0.91	--

Pipeline Diameter, Capital Cost, and Levelized Cost as a Function of Pipeline Length for Different CO2 Mass Flow Rates

Pipeline Length (km)	Key Output	Model						
		Ogden	*MIT*	*Ecofys*	*IEA GHG PH4/6*	*IEA GHG 2005/3*	*IEA GHG 2005/2*	*Parker*
CO2 Mass Flow Rate = 1000 tonnes/day								
100	Pipeline Diameter (in)	4.67	5.04	5.13	6.29	5.93	3.59	5.11
	Pipeline Capital Cost ($/km)	201,941	125,437	217,078	138,925	236,698	140,081	221,547
	Levelized Cost ($/tonne CO2)	10.89	6.31	10.29	6.26	11.14	7.07	--
200	Pipeline Diameter (in)	5.34	5.79	5.89	7.23	5.93	3.59	5.63
	Pipeline Capital Cost ($/km)	238,491	144,276	249,357	136,268	220,286	172,174	226,444
	Levelized Cost ($/tonne CO2)	25.72	14.13	23.65	12.46	20.74	18.10	--
300	Pipeline Diameter (in)	5.77	6.29	6.39	7.84	5.93	3.59	5.97
	Pipeline Capital Cost ($/km)	262,866	156,591	270,421	139,169	214,815	167,142	230,422
	Levelized Cost ($/tonne CO2)	42.52	22.68	38.47	19.05	30.34	26.39	--
400	Pipeline Diameter (in)	6.10	6.67	6.77	8.31	5.93	3.59	6.23
	Pipeline Capital Cost ($/km)	281,656	165,966	286,436	142,793	212,080	164,626	233,713
	Levelized Cost ($/tonne CO2)	60.75	31.74	54.33	25.98	39.93	34.68	--
500	Pipeline Diameter (in)	6.37	6.97	7.08	8.68	5.93	3.59	6.44
	Pipeline Capital Cost ($/km)	297,151	173,625	299,509	146,415	210,439	163,116	236,487
	Levelized Cost ($/tonne CO2)	80.12	41.22	71.01	33.18	49.53	42.97	--
CO2 Mass Flow Rate = 5,000 tonnes/day								
100	Pipeline Diameter (in)	8.70	9.53	9.77	11.98	13.27	8.04	10.21
	Pipeline Capital Cost ($/km)	437,251	237,340	413,241	216,933	529,273	189,985	300,110
	Levelized Cost ($/tonne CO2)	4.72	2.16	3.92	2.06	4.98	1.92	--
200	Pipeline Diameter (in)	9.94	10.96	11.22	13.76	13.27	8.04	11.20
	Pipeline Capital Cost ($/km)	516,390	272,800	474,690	237,516	492,575	222,081	316,200
	Levelized Cost ($/tonne CO2)	11.14	4.89	9.00	4.33	9.28	4.68	--
300	Pipeline Diameter (in)	10.75	11.89	12.17	14.93	13.27	8.04	11.84
	Pipeline Capital Cost ($/km)	569,167	295,961	514,788	257,206	480,342	217,050	327,716
	Levelized Cost ($/tonne CO2)	18.42	7.90	14.65	6.89	13.57	6.87	--
400	Pipeline Diameter (in)	11.37	12.59	12.89	15.81	13.27	8.04	12.33
	Pipeline Capital Cost ($/km)	609,853	313,582	545,275	274,454	474,226	214,534	336,970
	Levelized Cost ($/tonne CO2)	26.31	11.10	20.68	9.67	17.86	9.06	--
500	Pipeline Diameter (in)	11.87	13.17	13.48	16.53	13.27	8.04	12.73
	Pipeline Capital Cost ($/km)	643,403	327,972	570,161	289,746	470,556	213,024	344,764
	Levelized Cost ($/tonne CO2)	34.70	14.45	27.04	12.65	22.15	11.25	--
CO2 Mass Flow Rate = 10,000 tonnes/day								
100	Pipeline Diameter (in)	11.37	12.56	12.89	15.81	18.76	11.37	13.79
	Pipeline Capital Cost ($/km)	609,853	312,707	545,275	293,310	748,505	247,721	369,864
	Levelized Cost ($/tonne CO2)	3.29	1.38	2.59	1.40	3.52	1.25	--
200	Pipeline Diameter (in)	13.00	14.43	14.81	18.16	18.76	11.37	15.09
	Pipeline Capital Cost ($/km)	720,232	359,364	626,357	337,165	696,606	279,821	396,181
	Levelized Cost ($/tonne CO2)	7.77	3.14	5.94	3.04	6.56	2.96	--
300	Pipeline Diameter (in)	14.07	15.66	16.06	19.70	18.76	11.37	15.93
	Pipeline Capital Cost ($/km)	793,842	389,829	679,266	373,703	679,306	274,791	414,654
	Levelized Cost ($/tonne CO2)	12.84	5.08	9.66	4.93	9.59	4.36	--
400	Pipeline Diameter (in)	14.87	16.59	17.01	20.86	18.76	11.37	16.58
	Pipeline Capital Cost ($/km)	850,588	413,004	719,495	404,639	670,656	272,276	429,587
	Levelized Cost ($/tonne CO2)	18.35	7.15	13.65	7.01	12.63	5.76	--
500	Pipeline Diameter (in)	15.53	17.35	17.78	21.82	18.76	11.37	17.10
	Pipeline Capital Cost ($/km)	897,383	431,927	752,332	431,668	665,466	270,767	441,885
	Levelized Cost ($/tonne CO2)	24.20	9.32	17.84	9.27	15.66	7.17	--
CO2 Mass Flow Rate = 20,000 tonnes/day								
100	Pipeline Diameter (in)	14.87	16.56	17.01	20.86	26.53	16.07	18.65
	Pipeline Capital Cost ($/km)	850,588	412,186	719,495	423,483	1,058,545	359,124	483,837
	Levelized Cost ($/tonne CO2)	2.29	0.89	1.71	1.00	2.49	0.91	--
200	Pipeline Diameter (in)	17.01	19.02	19.54	23.97	26.53	16.07	20.36
	Pipeline Capital Cost ($/km)	1,004,538	473,624	826,483	507,440	985,149	391,232	527,225
	Levelized Cost ($/tonne CO2)	5.42	2.03	3.92	2.25	4.64	2.08	--
300	Pipeline Diameter (in)	18.40	20.63	21.19	25.99	26.53	16.07	21.47
	Pipeline Capital Cost ($/km)	1,107,206	513,734	896,297	573,040	960,684	386,204	557,490
	Levelized Cost ($/tonne CO2)	8.96	3.29	6.38	3.72	6.78	3.07	--
400	Pipeline Diameter (in)	19.45	21.86	22.44	27.53	26.53	16.07	22.32
	Pipeline Capital Cost ($/km)	1,186,353	544,241	949,379	627,600	948,451	383,690	581,626
	Levelized Cost ($/tonne CO2)	12.79	4.63	9.00	5.35	8.93	4.07	--
500	Pipeline Diameter (in)	20.31	22.86	23.46	28.79	26.53	16.07	23.01
	Pipeline Capital Cost ($/km)	1,251,619	569,149	992,709	674,894	941,111	382,182	601,788
	Levelized Cost ($/tonne CO2)	16.87	6.04	11.77	7.13	11.08	5.07	--

Pipeline Diameter, Capital Cost, and Levelized Cost as a Function of CO2 Mass Flow Rate (Common Diameters Applied, Pipeline Length = 100 km)

CO2 Mass Flow Rate (tonnes/day)	Key Output	Model						
		Ogden	*MIT*	*Ecofys*	*IEA GHG PH4/6*	*IEA GHG 2005/3*	*IEA GHG 2005/2*	*Parker*
1,000	Pipeline Diameter (in)	5.11	5.11	5.11	5.11	5.11	5.11	5.11
	Pipeline Capital Cost ($/km)	201,941	127,229	216,187	128,019	236,698	153,617	221,547
	Levelized Cost ($/tonne CO2)	10.89	6.38	10.25	5.64	11.14	7.76	--
2,000	Pipeline Diameter (in)	6.88	6.88	6.88	6.88	6.88	6.88	6.88
	Pipeline Capital Cost ($/km)	281,656	171,298	291,070	145,006	334,741	173,996	246,043
	Levelized Cost ($/tonne CO2)	7.59	4.07	6.90	3.30	7.88	4.39	--
4,000	Pipeline Diameter (in)	9.27	9.27	9.27	9.27	9.27	9.27	9.27
	Pipeline Capital Cost ($/km)	392,838	230,804	392,182	174,459	473,396	209,332	283,792
	Levelized Cost ($/tonne CO2)	5.30	2.64	4.65	2.04	5.57	2.64	--
6,000	Pipeline Diameter (in)	11.05	11.05	11.05	11.05	11.05	11.05	11.05
	Pipeline Capital Cost ($/km)	477,241	275,123	467,488	201,260	579,789	241,484	315,395
	Levelized Cost ($/tonne CO2)	4.29	2.05	3.69	1.59	4.55	2.03	--
8,000	Pipeline Diameter (in)	12.52	12.52	12.52	12.52	12.52	12.52	12.52
	Pipeline Capital Cost ($/km)	547,909	311,723	529,679	226,524	669,483	271,794	343,740
	Levelized Cost ($/tonne CO2)	3.69	1.72	3.14	1.35	3.94	1.72	--
10,000	Pipeline Diameter (in)	13.79	13.79	13.79	13.79	13.79	13.79	13.79
	Pipeline Capital Cost ($/km)	609,853	343,343	583,408	250,632	748,505	300,715	369,864
	Levelized Cost ($/tonne CO2)	3.29	1.51	2.77	1.20	3.52	1.52	--
12,000	Pipeline Diameter (in)	14.93	14.93	14.93	14.93	14.93	14.93	14.93
	Pipeline Capital Cost ($/km)	665,629	371,727	631,638	274,072	819,946	328,836	394,605
	Levelized Cost ($/tonne CO2)	2.99	1.35	2.50	1.09	3.22	1.38	--
14,000	Pipeline Diameter (in)	15.97	15.97	15.97	15.97	15.97	15.97	15.97
	Pipeline Capital Cost ($/km)	716,748	397,621	675,637	296,942	885,643	356,273	418,241
	Levelized Cost ($/tonne CO2)	2.76	1.23	2.29	1.01	2.98	1.29	--
16,000	Pipeline Diameter (in)	16.92	16.92	16.92	16.92	16.92	16.92	16.92
	Pipeline Capital Cost ($/km)	764,192	421,274	715,828	319,072	946,792	382,821	440,721
	Levelized Cost ($/tonne CO2)	2.58	1.14	2.12	0.95	2.79	1.21	--
18,000	Pipeline Diameter (in)	17.81	17.81	17.81	17.81	17.81	17.81	17.81
	Pipeline Capital Cost ($/km)	808,641	443,433	753,481	340,877	1,004,224	408,981	462,551
	Levelized Cost ($/tonne CO2)	2.42	1.06	1.98	0.90	2.63	1.15	--
20,000	Pipeline Diameter (in)	18.65	18.65	18.65	18.65	18.65	18.65	18.65
	Pipeline Capital Cost ($/km)	850,588	464,347	789,019	362,410	1,058,545	434,813	483,837
	Levelized Cost ($/tonne CO2)	2.29	1.00	1.87	0.86	2.49	1.10	--