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# Demonstrating $CO_2$ capture in the UK cement, chemicals, iron and steel and oil refining sectors by 2025: A Techno-economic Study

Final Report Appendix
for
**DECC and BIS**

30th April 2014

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**PSE Ltd**
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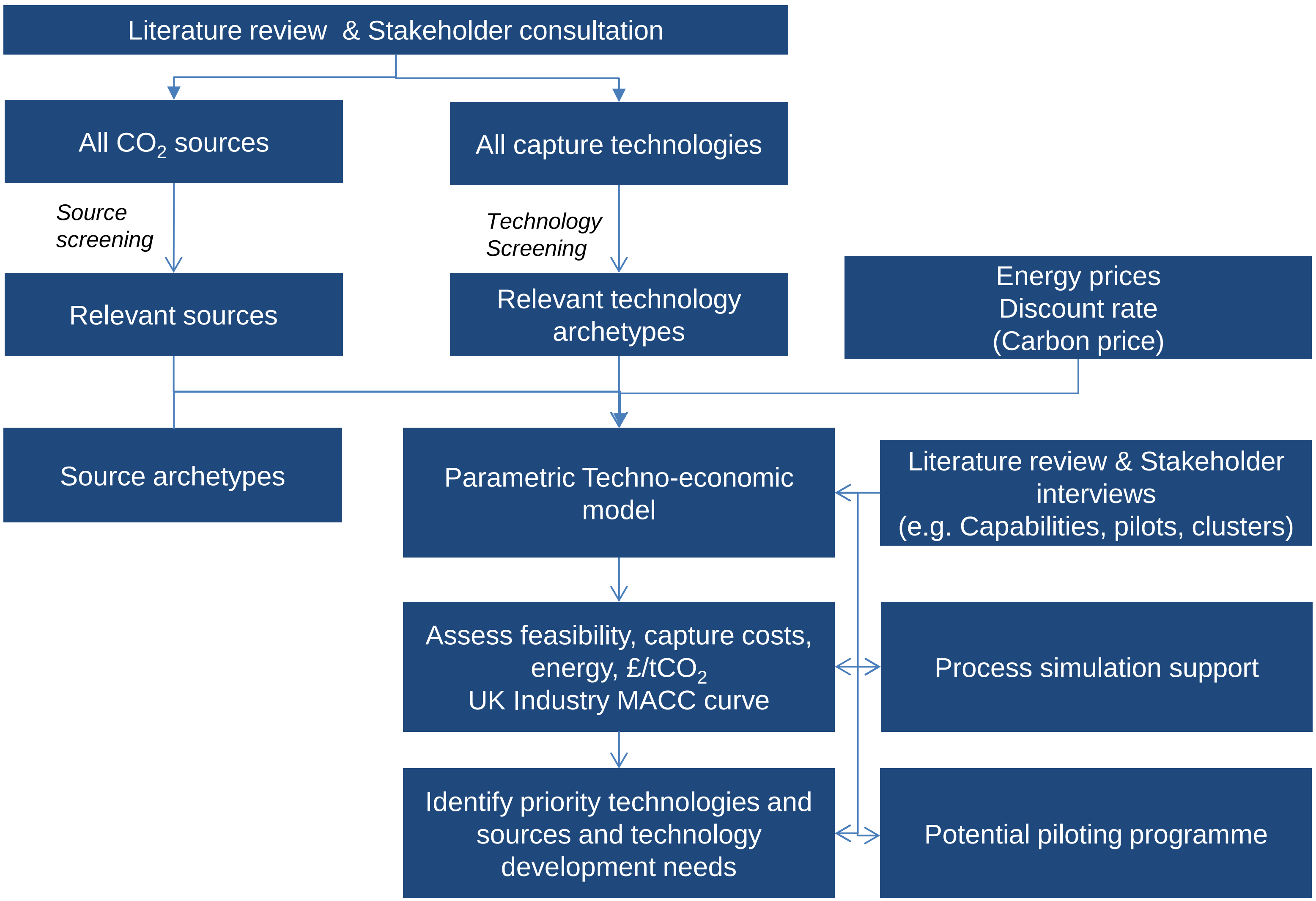
**University of Sheffield**


www.element-energy.co.uk

# Outline

- Overall Project Methodology
- $CO_2$ capture technologies
- $CO_2$ sources
- Techno-economic analysis of industrial $CO_2$ capture
- Process simulation case studies
- $CO_2$ utilisation review

# Overall project approach



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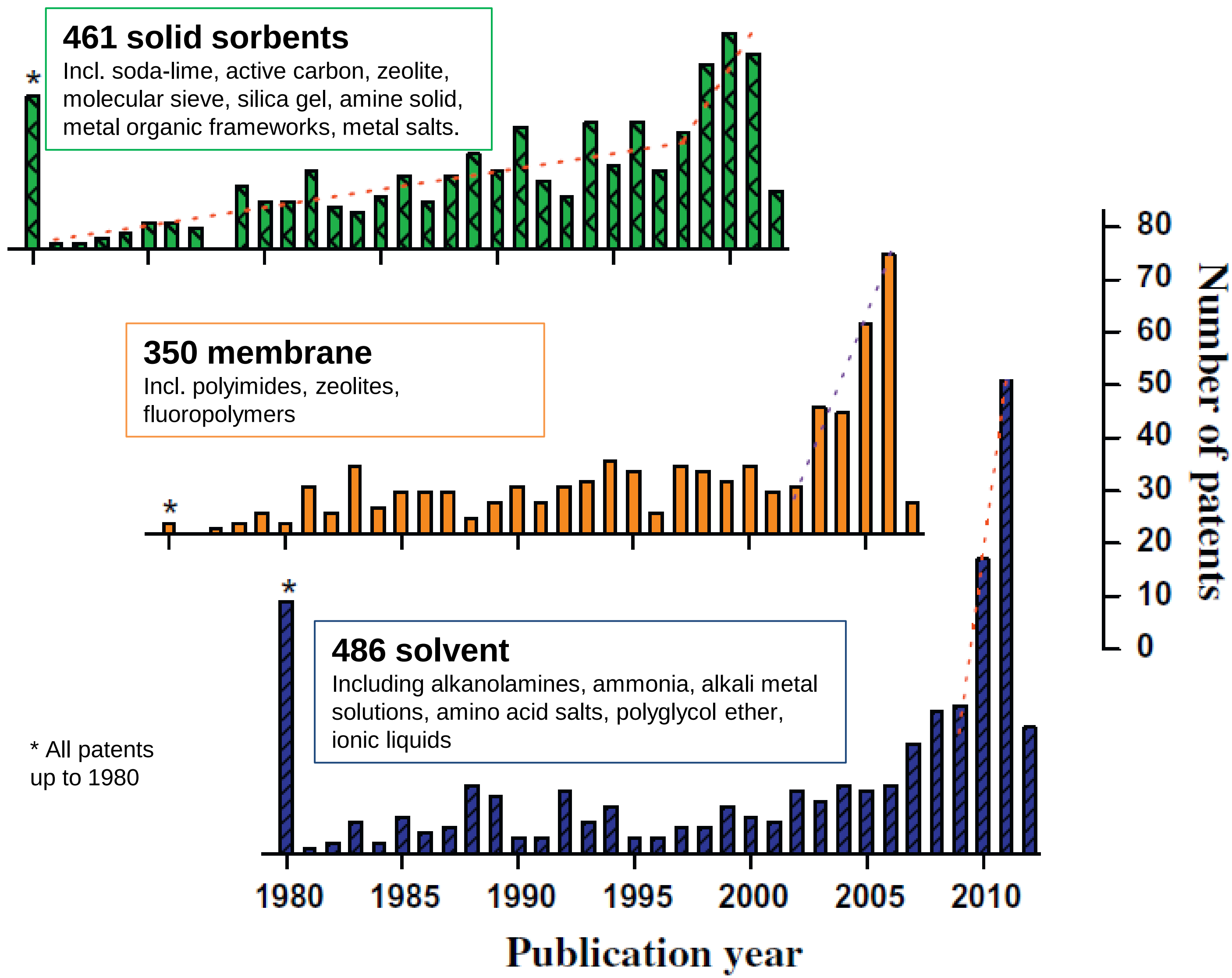
# Extensive literature review used to identify and characterise capture technologies

| Process | Key |
|---|---|
| Identify key data sources | • **Data search** using ScienceDirect, internet, WebofKnowledge, CCS portals, conference proceedings, manufacturer and CCS project websites. |
| Screening of data sources | • **Screening of abstracts of 2,000+ papers** to identify most relevant sources. |
| In-depth literature review | • **200+ papers reviewed in detail** for information across all known capture technologies |
| Filtering of technologies | • Review of agreed filters following literature review<br>• Initial shortlist reviewed by Imperial College and University of Sheffield |
| Populate capture technology database | • Techno-economic model design clarifies technology data collection requirements<br>• Focussed literature review **to populate database** |
| Review with stakeholders | • **Technology database circulated** with industry experts for review of assumptions |

# The pace of technology development: A recent review using identified *ca*. 1,300 patents on capture technologies.



# Resources to track global capture technology progress

Standard research engines (Google, Googlescholar, WebofKnowledge, ScienceDirect) were used to identify public literature for this project.

The most consistently useful and largely up-to-date curated information on pilots, demonstration and commercial scale CCS projects were found at the following portals:

http://www.globalccsinstitute.com/projects/browse

http://www.sccs.org.uk/news/2014/10Apr-GlobalCCSMap.html

http://www.cslforum.org/projects/index.html?cid=nav_projects

http://ieaghg.org/ccs-resources/rd-database

https://sequestration.mit.edu/tools/projects/index.html

http://www.zeroemissionsplatform.eu/

http://bellona.org/ccs/ccs-projects.html

These portals provide links to actual project websites, and facilitate understanding of the relative maturity (TRL) of each capture technology and sector. However some caveats are helpful: Published project factsheets generally provide only very basic data (location, scale, year, fundrs/partners). Conference reports offer latest findings, but rarely details. Peer reviewed academic papers in high quality journals generally describe the most successful results from pilots, but may lag years behind results, and few report pilot/demo costs and practical implementation challenges. As most pilots and demonstrations are co-funded by multiple partners and there are reporting differences in whether capital, total, marginal or average costs are paid for, it is challenging to compare the overall costs of pilots for different technologies, sometimes even to within one order of magnitude.

# Objective filtering process to identify most relevant capture technologies for techno-economic study

| Filter | Impacts |
| --- | --- |
| **All technology families** | • Capture technology long list identified |
| **Exclude TRL ≤4** | • Exclude technologies that have not been validated at lab bench scale. |
| **Exclude techs requiring base process re-design** | • Excludes oxycombustion, pre-combustion and several chemical looping, iron and steel (e.g. TGRBF and HIsarna) CCS project designs. |
| **Tech readiness** | • Assume operational plant for 0.05-5 Mt/yr by *ca*. 2025 requires 1000s of hours of successful operation at 0.01-1 Mt/yr respectively validated ahead of FID in *ca*. 2020. [FID assumed 2015 for operation in 2020].<br>• Excludes adsorption technologies, membranes, ionic liquids, hybrids. |
| **Insufficient data** | • Excludes carbonates, sodium hydroxide, and purisol based solvents |
| **Shortlist of 7 capture technologies** | • Chemical solvents: 1st gen amine, 2nd gen chemical solvents (e.g. amines, amino acids, and blends, potassium carbonate, ammonia)<br>• Physical solvents: rectisol, selexol<br>• Chemical looping: calcium looping<br>• Cryogenics: liquefaction |

# An initial "technology long list" comprised all capture technology families identified in the public literature.

A long list of technology families for $CO_2$ separation was identified from the CCS literature, and comprises:

**Liquid absorption** technologies**:**

- Chemical solvents, e.g. amines, ammonia, potassium carbonate solutions, other alkalis
- Physical solvents, e.g. rectisol, selexol, purisol, propylene carbonate, carbonates, ionic liquids

**Solid looping cycles**: e.g. calcium looping

**Adsorption** technologies

- Adsorber beds, e.g. alumina, zeolites, activated carbon
- Regenerative methods (temperature, pressure/vacuum, electrical swing adsorption, washing)

**Cryogenics** (i.e. purification by liquefaction or desublimation of $CO_2$)

**Membranes**

- Gas separation or absorption, Ceramics, amine-functionalised membranes

**Other** e.g. algae, enzymes, oxyfuel, pre-combustion, hybrid approaches

# Several capture technologies are likely to be available for retrofit bolt-on to industrial plants by 2025 and so included in the techno-economic modelling.

| $CO_2$ separation technology mechanism | Technology | Criteria | | | Include in techno-economics? |
|---|---|---|---|---|---|
| | | Technology availability (TRL ≥5?) | Feasibility as bolt-on? | Sufficient data available? | |
| Chemical absorption | Amines, first generation (MEA, MDEA, KS-1) (with temperature swing) | 9 (nat gas /high purity)<br>7-8 (power and industrial) | Bolt on | Yes | Yes |
| | Amines, second generation | 6-7 | Bolt on | Yes | Yes |
| | Ammonia | 6-7 | Bolt on | Yes | Yes |
| | Potassium carbonate solution (with pressure swing, amine promoted) | (9 Fischer Tropsch)<br>7 Flue gas | Bolt on | Yes | Yes |
| | Alkalis | 6 | Bolt on | No | No |
| Physical absorption | Rectisol (methanol) | 9 (nat gas/syngas/high purity) | Bolt on | Yes | Yes |
| | Selexol (glycol) | 9 (nat gas/syngas/high purity) | Bolt on | Yes | Yes |
| | Purisol | 9 (nat gas/syngas/high purity) | Bolt on | No | No |
| | Propylene carbonate | 9 (nat gas/syngas/high purity) | Bolt on | No | No |
| | Dimethyl Carbonate + analogues diethyl carbonate etc. | 4-5 | Bolt on | No | No |
| | Ionic liquids (can be chemical abs) | 4-5 | Bolt on | No | No |
| Solid looping cycles | Calcium looping | 6-7 | Bolt on and redesign options | Yes | Yes |
| Cryogenics | Cryogenics (i.e. purification by liquefaction or desublimation of $CO_2$) | 9 (high purity only)<br>6 (flue gases) | Yes | Yes | Yes |

Ref: Olajire (2010) *Energy* **35** 2610-2628

# Some $CO_2$ capture technology families are not included in the techno-economic modelling as they involve significant plant redesign or insufficient data are available.

| $CO_2$ separation technology mechanism | Technology | Criteria | | | Include in techno-economics? |
|---|---|---|---|---|---|
| | | Technology availability (TRL ≥5?) | Feasibility as bolt-on? | Sufficient data available? | |
| Membranes | Gas separation (e.g. polyphenyleneoxide, polydimethylsiloxane) or absorption (polypropylene), Ceramics, amine-functionalised membranes | 8-9 (highest for nat. gas, syngas, high purity)<br>5-6 (low purity flue gas) | Yes | No | No |
| Adsorption | Adsorber beds, e.g. alumina, zeolites, activated carbon | 5 (flue gas) | Bolt on | No | No |
| Adsorption | Regenerative methods (temperature, pressure/vacuum, and electrical swing adsorption). | 5 (flue gas) | Bolt on | No | No |
| Biological | Algae | 4 | Yes | No | No |
| Biochemical | Enzymatic conversion or enzyme promoted reaction | 4 | Yes | No | No |
| Oxyfuel | Oxyfuel | 7-8 (power generation)<br>6-7 (industrial applications) | Involves significant plant redesign | Some applications | No |
| Pre-combustion | Pre-combustion | 8-9 (syngas stream in IGCC or SMR)<br>6-7 (other gasification processes for industrial applications) | Involves significant plant redesign | No | No |
| Mixed | Hybrid technologies (e.g. combination of solvents, PSA, TSA, cryogenics, membranes, recirculation, compression) | 4-6 (wide range of options) | In some cases | Some applications | No |

# $CO_2$ capture technologies

- Archetype data identified in technology database
- Chemical absorption solvents
  - 1st generation amine
  - 2nd generation amine
  - Potassium carbonate
  - Ammonia
- Physical absorption solvents
  - Rectisol
  - Selexol
- Solid Looping
  - Calcium looping
- Cryogenic capture

# The technology database lists key attributes to be used in the techno-economic modelling.

Capture technology database lists:

1. Technology name and family
2. Current TRL and commercial availability for operation in 2013, 2020 and 2025
3. Reference input *and* output conditions for temperature, pressure, and impurity composition
4. $CO_2$ input (size, % mol fraction) reference conditions for capex
5. Capex of reference project (and uncertainty).
6. Relative cost in 2020 and 2025
7. Annual opex as a % of capex (current, 2020, 2025)
8. Average GJ thermal energy required/t$CO_2$ captured (current, 2020, 2025)
9. MWh Electrical energy needed/t$CO_2$ captured
10. Space required (low/high)
11. Feasibility for retrofit bolt-on (Y/N)
12. Cooling water required (low/high)
13. Process water required (low/high)
14. COMAH status required at site (top/low/none)
15. Complex process requiring skilled workers (Y/N)
16. Pre-development + construction period (only specified if not default of 3+3 yrs)

# Baseline modelling assumptions for capture technology archetypes

| Technology Name | Approx TRL | Minimum input overall $CO_2$ stream pressure (MPa) | Impurity tolerance (ppm) | | Output $CO_2$ (% volume) | Output $CO_2$ stream pressure (MPa) | Reference capex (£m 2013) (+100%/-50%) | Reference $CO_2$ (Mt captured/y) | Reference $CO_2$ purity (% volume) | Fixed opex (% of capex in 2013) | Thermal GJ/t$CO_2$ captured | | Electrical GJ/t$CO_2$ captured | | Relative capex (2013 = 100%) | | Relative opex (2013 = 100%) | | Capture efficiency (amount captured/ input $CO_2$) | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | NOx | SOx | | | Central | | | Central | 2020 | 2025 | 2020 | 2025 | 2020 | 2025 | 2020 | 2025 | 2020 | 2025 |
| 1st gen amine | 8 | 0.1 | 10 | 10 | 99.0% | 0.1 | £ 462 | 2 | 11.5% | 8% | 3.8 | 3.6 | 0.2 | 0.2 | 100% | 90% | 80% | 60% | 85% | 90% |
| Advanced amines or blends | 7 | 0.1 | 100 | 100 | 99.0% | 0.1 | £ 355 | 2 | 11.5% | 5% | | 3. 0 | | 0.2 | 100% | 77% | 80% | 60% | 85% | 90% |
| Chilled ammonia | 7 | 0.1 | 10 | 10 | 99.9% | 0.1 | £ 380 | 2 | 11.5% | 8% | | 3.0 | | 0.6 | 100% | 100% | 80% | 60% | 85% | 90% |
| Potassium carbonate | 9 | 3.0 | 200 | 200 | 90.5% | 0.1 | £ 399 | 2 | 5.0% | 7% | 5.0 | 5.0 | 0.5 | 0.5 | 100% | 100% | 80% | 60% | 85% | 90% |
| Rectisol | 9 | 3.0 | 100 | 100 | 98.5% | 0.1 | £ 200 | 2 | 35.0% | 5% | 0.4 | 0.4 | 0.2 | 0.2 | 100% | 100% | 80% | 60% | 85% | 90% |
| Selexol | 9 | 3.0 | 100 | 100 | 99.0% | 0.1 | £ 190 | 2 | 40.0% | 5% | 0.2 | 0.2 | 0.2 | 0.2 | 100% | 100% | 80% | 80% | 85% | 90% |
| Calcium looping | 6 | 0.1 | 100 | 100 | 90.0% | 0.1 | £ 142 | 2 | 13.0% | 19% | | 1.6 | | 0.54 | | 100% | | 80% | | 85% |
| Cryogenics | 7 | 0.1 | 10 | 10 | 99.0% | Liquid CO2 | £ 290 | 2 | 13.5% | 5% | - | - | | 3.6 | 100% | 100% | 80% | 80% | 85% | 90% |

- Capture plant output pressures are assumed to be 1 bar, for all technologies.

# $CO_2$ capture technologies

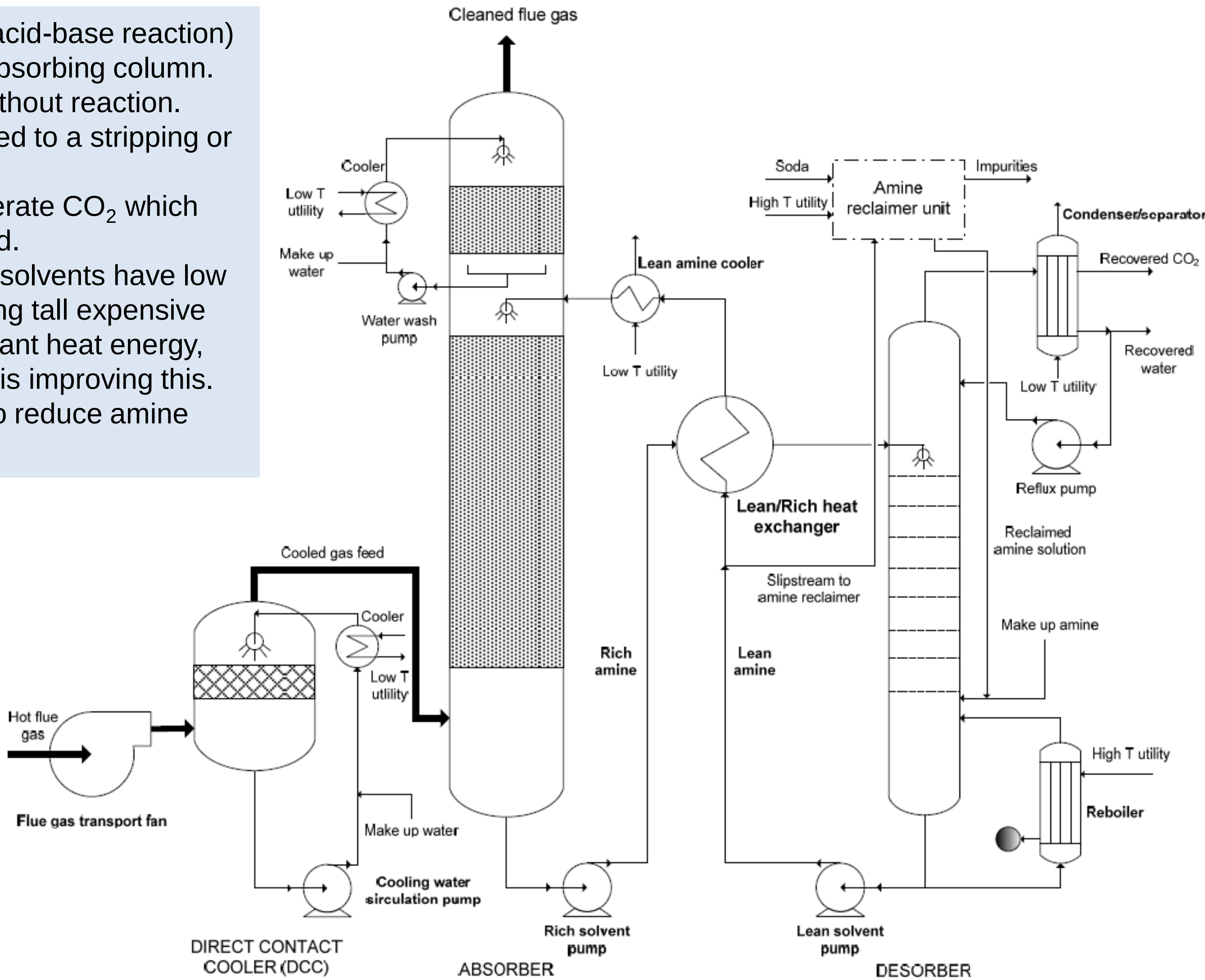
- Archetype data identified in technology database
- Chemical absorption solvents
  - 1st generation amine
  - 2nd generation amine
  - Potassium carbonate
  - Ammonia
- Physical absorption solvents
  - Rectisol
  - Selexol
- Solid Looping
  - Calcium looping
- Cryogenic capture

# First generation amine solvent (e.g. monoethanolamine MEA)

- **Brief description**- MEA selectively absorbs $CO_2$ from flue gas (temperatures between 40 and 60°C), and is then sent to the striper where $CO_2$-rich MEA solution is heated (100–140°C at atmospheric pressure) to release almost pure $CO_2$. The $CO_2$-lean MEA solution is then recycled to the absorber.
- **Technology status**- MEA is the most widely used solvent for $CO_2$ capture. Technology developed over 70 years ago to remove acid gasses from natural gas streams and has been currently being optimised for flue gas $CO_2$ capture. Commercially available at 0.1MtCO2/yr scale to produce high purity $CO_2$ for the food industry and acid gas sweetening. R&D is being done to target sector specific flue gases (e.g. cement industry), decrease corrosion and desorption process improvements.
- **Technology providers**- Fluor (Econamine FG), ABB/Lummus, Mitsubishi (KM-CDR), HTC Purenergy, Aker Clean Carbon, Cansolv (Absorbent DC101)
- **Economic and market factors**- Works well with low partial pressure and mild temperature flue gas. Well understood technology, already implemented in large scale projects. High recovery rates and purity.
- **Key barriers and challenges**- High energy requirements due to solvent regeneration, solvent degradation and equipment corrosion, environmental impacts due to solvent emissions and large absorber volume. High capex due to low $CO_2$ loading resulting in large absorber volume

# The main steps in chemical absorption capture are absorption and desorption (stripping).

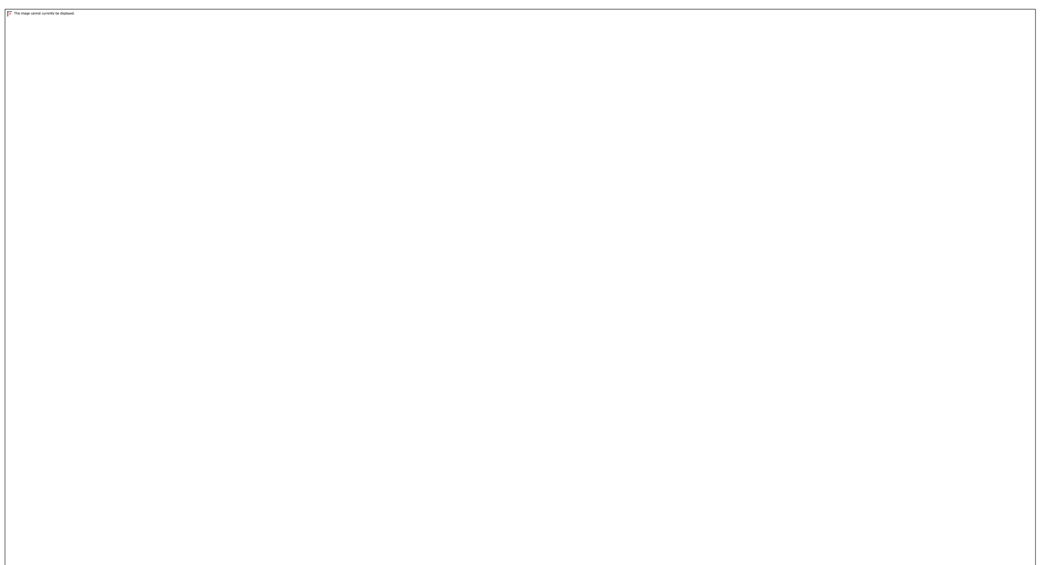
- $CO_2$ reacts chemically (e.g. acid-base reaction) with aqueous solvent in an absorbing column.
- Other gases pass through without reaction.
- The product is then transferred to a stripping or desorber column.
- Typically this is heated to liberate $CO_2$ which can be dried and compressed.
- Currently available chemical solvents have low $CO_2$ loading capacity (implying tall expensive columns) and require significant heat energy, but technology development is improving this.
- Equipment is also included to reduce amine loss to atmosphere.




OR STRIPPER

# Example reference project costs for a first-of-a-kind 1st generation amine post-combustion capture project: Longannet (1/2)

- After a thorough literature review, the most detailed and up-to-date publicly available cost study of 1st generation amine-based capture at scale in the UK is provided by the published Longannet CCS FEED study, produced by Scottish Power and partners for the first DECC CCS Competition.
- Note this represents a study that was not developed to completion, i.e. the costs are estimated rather than realised. Nevertheless the costs are detailed, adjusted for UK conditions, and represent a “first-of-a-kind” project conditions.
- This project considered retrofit post-combustion capture using Aker Clean Carbon’s amine-based solvent at the existing Longannet coal power station.
- Heat and power for the capture plant provided by a new gas boiler.



Scottish Power CCS Consortium (2011) UK DECC CCS Demonstration Competition SP6.0 RT015 FEED Close Out Report – CCS Project Costs

# Example reference project costs for a first-of-a-kind 1st generation amine post-combustion capture project: Longannet (2/2)

- Costs used for this study include SPS (£115m, steam and power supply), CCP (£228m, carbon capture plant) and BoP (£120m, balance of plant), which amount to £462m. An additional £122m (*ca*. 25%) is identified in the Longannet FEED study for risk/contingency, making a total of £584m for a plant of ca. 2Mt$CO_2$/yr, with a flue gas input stream concentration of 11.5%$CO_2$.
- This cost estimate excludes $CO_2$ compression.
- As the Longannet scenario includes initial pre-treatment to reduce $NO_X$ and $SO_X$ levels to less than 10 ppm, which is modelled separately in the techno-economic model here, an estimated cost of pre-treatment of £122m was deducted from this total, to give a Total Plant Cost estimate of £462m.
- This figure was used as the "reference" cost for a first-of-a-kind 2Mt$CO_2$/yr amine capture plant, and an uncertainty of +100%/-50% assumed.
- Although this estimate is at the upper range of published costs for 1st generation amine capture, this may be because the majority of published studies reference "nth of a kind" plant, or fail to specify conditions.

# Second generation chemical solvents

- **Brief description**- a large number of amines are being investigated worldwide to identify molecules with higher performance (e.g. faster reaction kinetics, higher $CO_2$ loading, lower heat requirement, lower environmental impacts through lower volatility/by-products, wider tolerance of conditions *cf.* first generation technologies MEA, MDEA etc.).
- **Technology status**- TRL6 the technologies draw on the processes optimised for first generation capture technologies, but will need re-optimising and demonstration under realistic operating conditions.
- **Technology providers**- tens of different technology developers worldwide (although some are inexperienced), examples include Mitsubishi, Aker, KEPCO Research Institute, Carbon Clean Solutions.
- **Economic and market factors**- lower energy costs associated with lower solvent thermal regeneration, lower material costs for solvents that are less corrosive than MEA, lower solvent costs for solvents with higher stability and reduced volatility. However, more "bespoke" molecules may be inherently more expensive and suffer from lack of economies of scale.
- **Key barriers and challenges**- conservative investors will need to see reference projects with thousands of run hours at similar flue gas conditions and scale. Not necessarily straightforward to "swap" amines in existing capture plants. Amine processes are likely to only be validated for a narrow range of pressure, temperature and impurity composition for flue gas (with reaction between amines and SOx and NOx that form salts). Some techs developed by academics or spin-outs without manufacturing capacity or clear route to market. Standard risks for immature techs.

# Wide range of 2nd generation chemical solvents are being developed, although it is not yet clear which will have the optimal properties.

As well as first generation solvents such as MEA and MDEA, second generation solvents under investigation are:

- Aminoethylethanolamine (AEEA)
- Piperazine (PZ) (used on own, or more commonly as a promoter)
- Tetraethylenepentamine (TEPA)
- Diethanolamine (DEA)
- Triethanolamine (TEA)
- 2-amino-2-methyl-1,3,propanediol (AMPD)
- Diisopropanolamine (DIPA)
- Polyethyleneimine (PEI)
- 3-aminopropyltriethoxysilane (APTES)
- Diethyltriamine (DETA)
- Diglycolamine (DGA)
- 2-amino-2-methylpropanol (AMP)
- Different concentrations of above
- Amino acids
- Mixtures “blends” of amines under intense investigation
- Amine/carbonate mixtures
- Wide range of temperature/pressure and process integration conditions under investigation
- Amine-functionalised adsorbent surfaces (zeolites, membranes)
- Proprietary solvents

# Assumptions for 2nd generation chemical solvent capture technologies

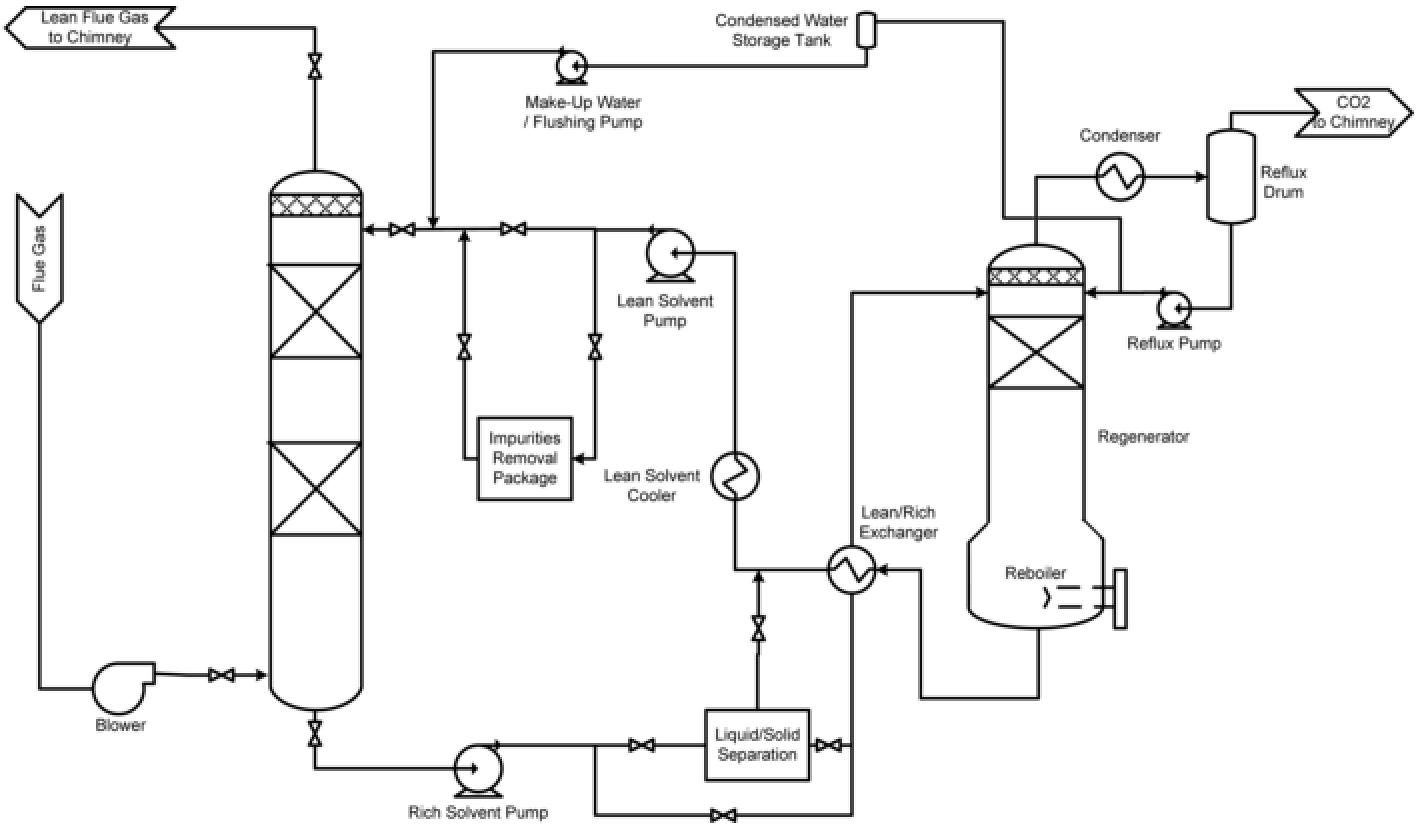
- Following literature review and discussions with technology developers, the assumed properties of a 2nd generation chemical solvent are developed by reference to 1st generation capture technologies as follows:
  - Capex 77% of 1st generation amine capture capex, based on reduced equipment sizes and less expensive alloys.
  - Fixed opex 5% of capex (cf. 8% for 1st generation amine), driven by lower solvent, water, environmental and waste disposal costs
  - Heat requirement 3 GJ/t$CO_2$ captured (cf. 3.6 GJ/t$CO_2$ captured for 1st generation amine).
  - No change in power demand (assume output of 1 bar $CO_2$)
  - Tolerance to 100 ppm NOx and 100 ppm SOx (compared to 10 ppm assumed for 1st generation amines).
  - Note these properties are based on extrapolation from lab-scale results; there is no guarantee that these will be fully realised for initial large-scale projects.

Values calculated assume the reductions identified by the DECC TCE CCSA Industry Cost Reduction Task Force Final report (page 27) on post-combustion capture (10%+13%=23%) can be achieved in time for a project operational by 2025. DoE NETL technology paper, conversations with tech developers at CCS conferences, and focussed interview with CCS Solutions advised of multiple opportunities for cost reduction, to reduce column sizes, substitute less expensive alloys, reduce redundancy, reduce clean up equipment, and reduced boiler costs as lower heat demand. Realistically not all of these can be delivered for a project operational by 2025. Choice of Longannet reference point likely to include appropriate risk premium for first of a kind technology implementation. Assumptions reviewed by Imperial College London, and CCS Solutions.

# Potassium Carbonate

- **Brief description**- Hot potassium carbonate absorbs $CO_2$ from flue gas in an absorber (potassium carbonate in the absorber is at c.a. 100°C and 10 bar). The CO2 rich solvent is then sent to a regenerator where the process is reversed by pressure reduction and heating. The $CO_2$-lean hot potassium carbonate solution is then recycled to the absorber. Activators and inhibitors are usually added to improve $CO_2$ absorption and to inhibit corrosion, and novel blends provide opportunities for future cost reduction or performance improvement.
- **Technology status**- Technology has been used to remove acid gases in a range of industrial processes, primarily synthetic gas, since it was first developed in the 1950s by Benson and Field. Benfields is the most common process, followed by CANTABARB.
- **Technology providers**- UOP (Benfields Process), Eickmeyer & Associates (CATABARB Process), Exxon (Flexsorb HP process)
- **Economic and market factors**- Works well with high partial pressure and mild temperature gases. Well understood technology, already implemented in large scale projects. High recovery rates and purity.
- **Key barriers and challenges**- High energy requirements due to solvent regeneration and high pressure required to operate. Equipment corrosion and environmental impacts can be a problem due to solvent emissions.
- **Capital cost estimates:** Because of different assumptions of majority of carbonate papers and amine papers, use a relative cost, and then estimate cost relative to 1st generation chemical amine solvent reference (e.g. Longannet). Rochelle, A.G.T. et al., 2007. $CO_2$ Capture by Absorption with Potassium Carbonate. Similar cost ratio MEA/Potassium carbonate obtained (288.3/352) from Oexmann, J., Hensel, C. & Kather, A., 2008. Post-combustion CO2-capture from coal-fired power plants: Preliminary evaluation of an integrated chemical absorption process with piperazine-promoted potassium carbonate. International Journal of Greenhouse Gas Control, 2(4), pp.539–552.

# Potassium carbonate (Benfield) process flow diagram

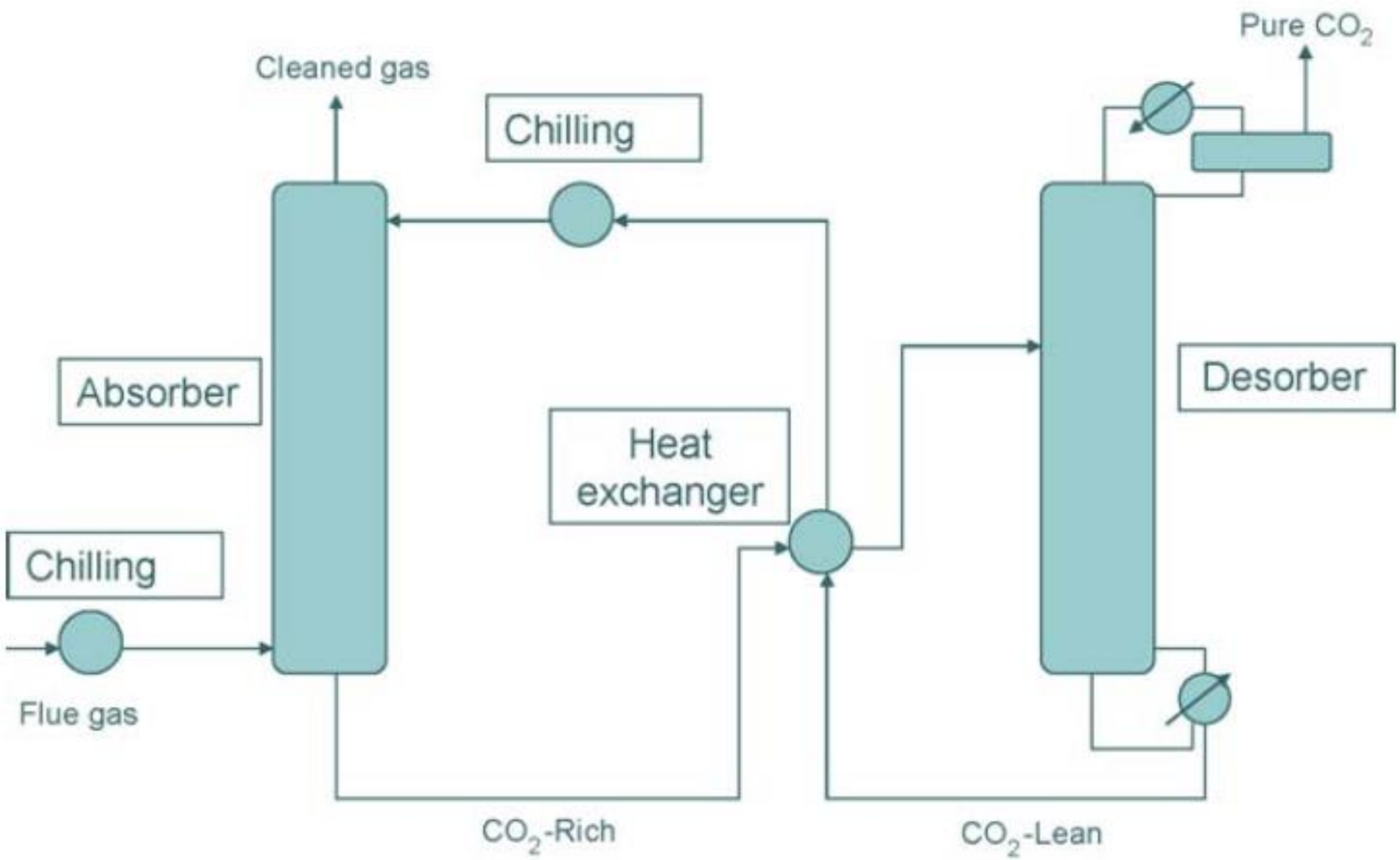



Source: Smith, K et all. (2014). Demonstration of a concentrated potassium carbonate process for CO2 capture. *Energy & Fuel*, *28*(1).

# Ammonia-based capture

- **Brief description**- temperature of the flue gas is reduced to 0-10°C to achieve maximum condensation and gas cleaning effect (removes practically all $SO_2$, $SO_3$, $NO_2$ and ash from the flue gas. Flue gas containing $CO_2$ is contacted with ammonium carbonate solution in water in an Absorber. The $CO_2$ in the flue gas reacts with the ammonia to form ammonium bicarbonate. The $CO_2$ rich solution is then pumped to a Desorber (or regenerator) where heat is applied for regeneration of the solution and release of $CO_2$. The ammonia solution is then returned to the absorber for reuse.
- **Technology status**- Piloting has been successful, and the technology has been proposed for pre-commercial projects, which are being planned.
- **Technology providers**- Alstom is marketing “chilled ammonia” (demonstrated at a power plant for 112,500 $tCO_2$/yr); NETL-Powerspan are marketing “aqueous ammonia”.
- **Economic and market factors**- Feasibility of multi-pollutants capture. High pressure regeneration reduces capital cost and energy consumption of the $CO_2$ capture plant relative to MEA. $CO_2$ uptake per kg of ammonia is estimated to be 3 times that per kg of MEA. High pressure $CO_2$ output could facilitate future $CO_2$ transport, although as there is limited public detailed data on energy/mass, cost, performance and pressure scenarios, assume 1 bar output for the techno-economic modelling.
- **Key barriers and challenges**- Ammonia is a volatile toxic gas, and storage and use of significant levels of ammonia may need a high COMAH status. The risk of accidental release of ammonia (slip) in the flue gas could be a significant concern. The main opportunities for cost reduction are through increased process integration.

# Chilled ammonia process flow diagram




Source: Darde, V. *et al*., 2010. Chilled ammonia process for $CO_2$ capture. *International Journal of Greenhouse Gas Control*, 4(2), pp.132.

# Assumptions for chilled ammonia capture

- In chilled ammonia capture processes, $CO_2$ can be stripped off at pressures up to 30 bar, potentially reducing the need for post-capture compression relative to MEA-based capture for which $CO_2$ is often modelled at *ca.* 1 bar prior to compression.
- However, for the high level techno-economic modelling we have assumed an output pressure as constant 1 bar for all the technologies.
- This is a conservative simplifying assumption, which has been made as details of mass and energy balances and costs including breakdown of compression requirements have not been well described for ammonia capture with industrial sources.
- Whilst most papers agree on the likely heat requirements for capture are likely to be less than that of 1st generation amine-based capture (a value of 3 GJ/t$CO_2$ captured for a coal reference source is assumed) papers disagree over relative capital and costs.
- The most recent, transparent and detailed reference public cost estimate for chilled ammonia capture has been prepared by Versteeg and Rubin (2011) who estimate a minimum cost of US$424m in $(2007), which when corrected for engineering cost inflation (IHS CERA US PCCI index excluding nuclear), UK location (factor 1.2) and then converted to GBP ($1.64/£1) gives £380m £(2013).

Versteeg and Rubin (2011) Technical and economic assessment of ammonia-based post-combustion $CO_2$ capture, *Energy Procedia* **4** 1957-1964

# References - Chemical Absorption (1/2)


- Abu-Zahra, M. R. M., Niederer, J. P. M., Feron, P. H. M., & Versteeg, G. F. (2007). CO2 capture from power plants. *International Journal of Greenhouse Gas Control*, *1*(2), 135–142. doi:10.1016/S1750-5836(07)00032-1
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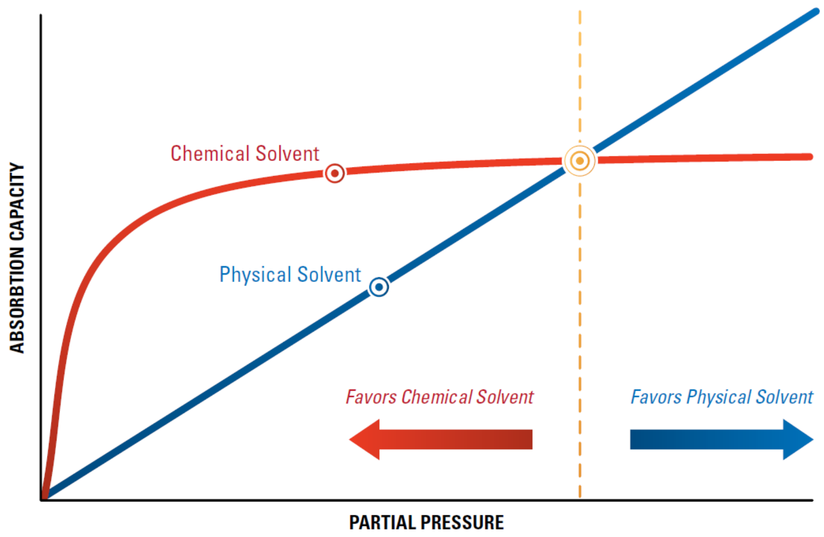
# $CO_2$ capture technologies

- Archetype data identified in technology database
- Chemical absorption solvents
  - 1st generation amine
  - 2nd generation amine
  - Potassium carbonate
  - Ammonia
- Physical absorption solvents
  - Rectisol
  - Selexol
- Solid Looping
  - Calcium looping
- Cryogenic capture

# Physical solvents can be applied at high partial pressures of $CO_2$ and with limited heat requirement.

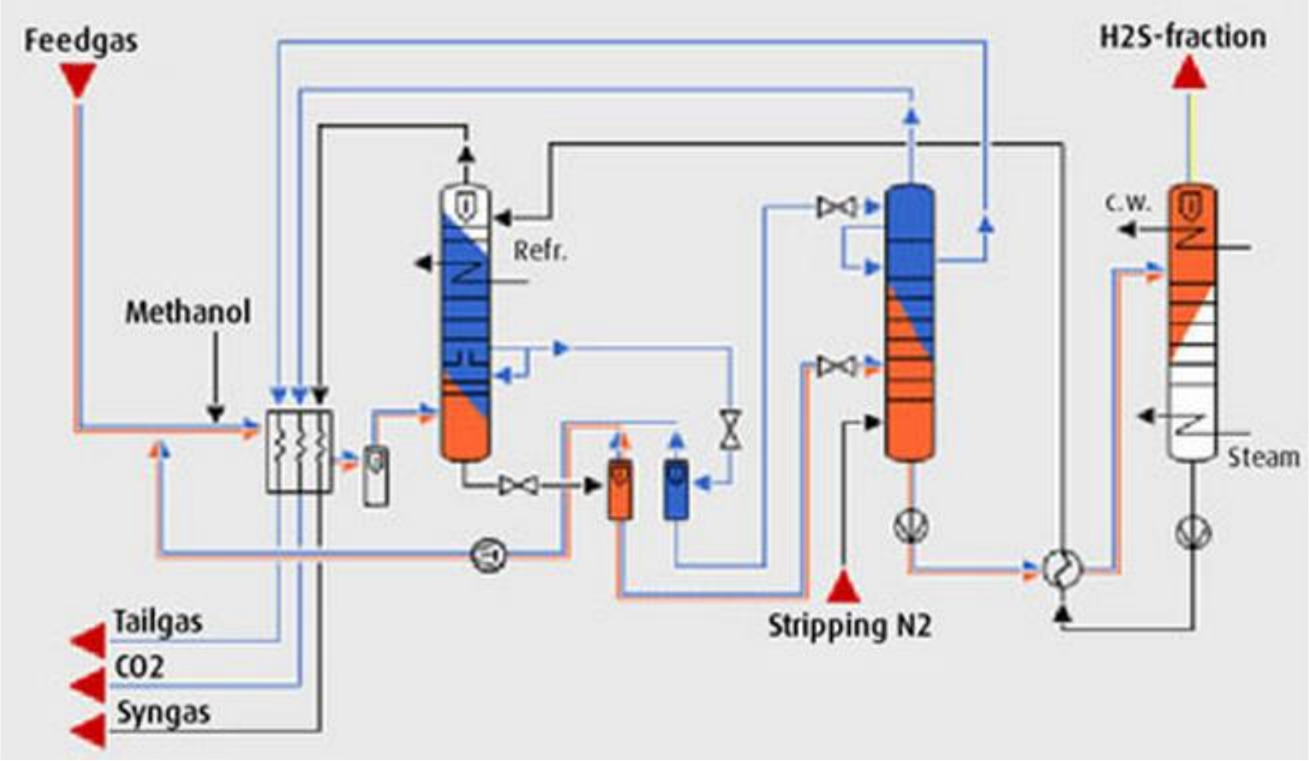
- Gaseous component dissolves into a liquid solvent (without reacting) forming a solution.
- Gases have different solubilities, so the solvent can be used selectively to separate the different gas components.
- Since $CO_2$ interacts weakly with physical solvents, these typically have low thermal regeneration energy demands. However $CO_2$ pressure is usually lost on release.
- The cost and performance (including selectivity) is dependent on scale, temperature, pressure and feedgas stream composition. FOAK premia for industrial CCS are not well understood.
- It is common to use these at low temperature, high $CO_2$ partial pressure, and with natural gas or syngas feedgas (rather than combustion flue gases).
- Output $CO_2$ pressure is variable but assumed 1 bar for the techno-economic modelling.



# Rectisol

- **Brief description**- $H_2S$ free flue gas is cooled to -20°C and fed to the absorber where the $CO_2$ dissolves in the cold methanol (kept at around -30°C). The $CO_2$ rich methanol is then fed to a flash drum where the $CO_2$ is separated. The methanol is then cooled and returned to the absorber for reuse.
- **Technology status**- Rectisol wash was developed in the 1950s, and is mainly used for sour gas purification. Commercial scale Rectisol units are operated world-wide for various processes (purification of hydrogen, production of ammonia, production of syngas for methanol synthesis, production of pure carbon monoxide and oxogases)
- **Technology providers**- Linde AG, Lurgi AG
- **Economic and market factors**- preferred for high pressure flue gas. Uses a cheap, low toxicity, low corrosion and easily available, non-proprietary solvent. It is flexible in process configuration. It can remove greater percentages of acid gas components providing a higher purity gas than other solvents. High $CO_2$-loading capacity allows for lower solvent flow rates compared to other physical solvent processes such as Selexol.
- **Key barriers and challenges**- Significant capital and operational costs are required. High energy usage for refrigeration and high vapour pressure of methanol causes solvent losses. Most cost effective with high pressure feed gasses since high pressure is required for flash drum operation which may increase operational cost with low-pressure feed gasses.
- **Capital cost estimate –** Based on 2012 cost estimates supplied by developer (confidential) and adjusted to 2013 to give a reference project capex of £200m.

# Rectisol Process flow diagram (combined $H_2S$ and $CO_2$ separation)

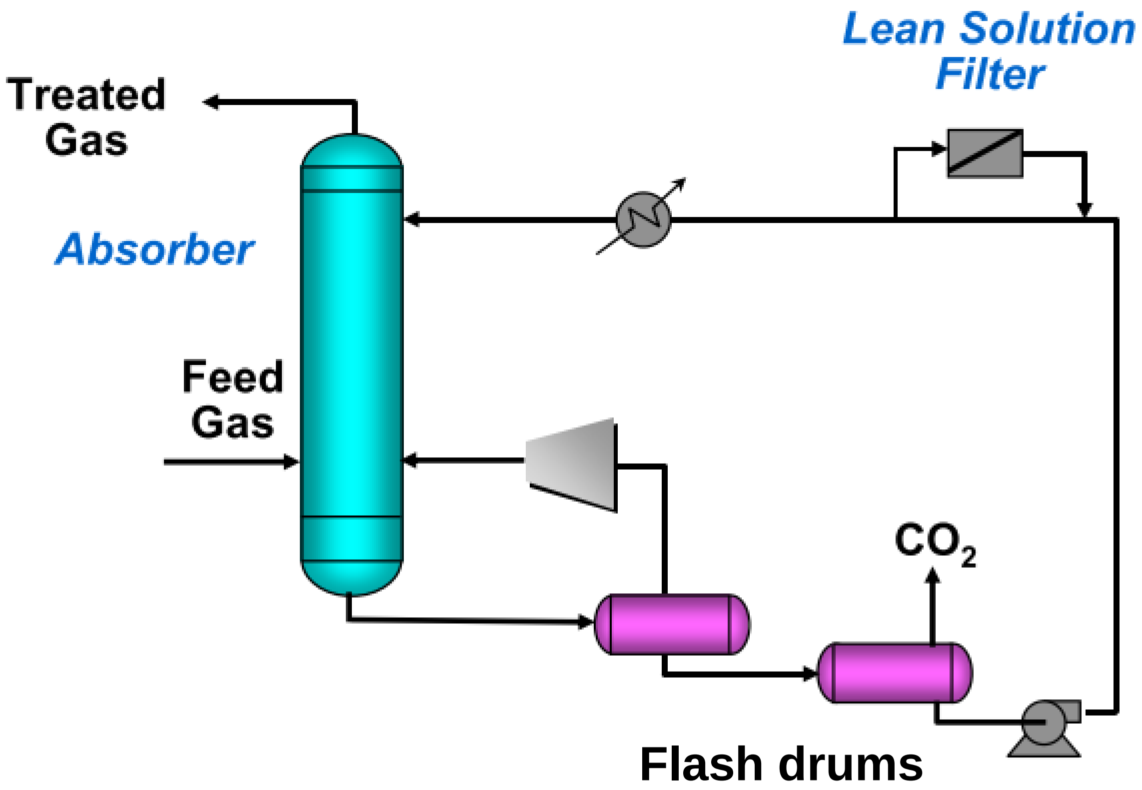



Source: http://www.linde-engineering.com

# Selexol

- **Brief description**- Flue gas (temperature around 35°C) is fed to the absorber where the $CO_2$ dissolves in the glycol. The $CO_2$ rich glycol is then fed to a series of flash drums where the $CO_2$ is separated. The glycol is then returned to the absorber for reuse.
- **Technology status**- Selexol is a licenced process that has been used commercially for 30 years and has over 60 units in commercial service, particularly for gas sweetening in the oil and gas industry. Specific to carbon capture, several commercial–scale carbon capture projects are under construction- mostly pre-combustion (Nuon Willem Alexander IGCC, Kemper County project, Green Gen).
- **Technology providers**- Dow, UOP (using Dow solvent), Clariant, Uhde GmbH
- **Economic and market factors**- particularly effective with high-pressure, low-temperature, flue gases. It has low toxicity and is a less corrosive solvent. Compared to amines, it has a higher capacity to absorb $CO_2$ at high pressure, requires less heat for solvent regeneration and $CO_2$ is delivered at higher pressures, meaning less compression is necessary for utilisation/transport or storage. It can operate selectively to capture different gases (e.g. hydrogen sulfide, carbon dioxide)
- **Key barriers and challenges**- Most cost effective with high pressure feed gasses since high pressure is required for flash drum operation which may increase operational cost with low-pressure feed gasses. Missing large-scale post combustion trials.
- **Cost estimates** – Costs derived from published IGCC-CCS scenarios in the NETL cost and performance baseline studies (e.g. $213m), and adjusted for UK location (factor 1.2), inflation and GBP ($1.64/£1), to give a reference project capex of £190m.

# Selexol process flow diagram



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# $CO_2$ capture technologies

- Archetype data identified in technology database
- Chemical absorption solvents
  - 1st generation amine
  - 2nd generation amine
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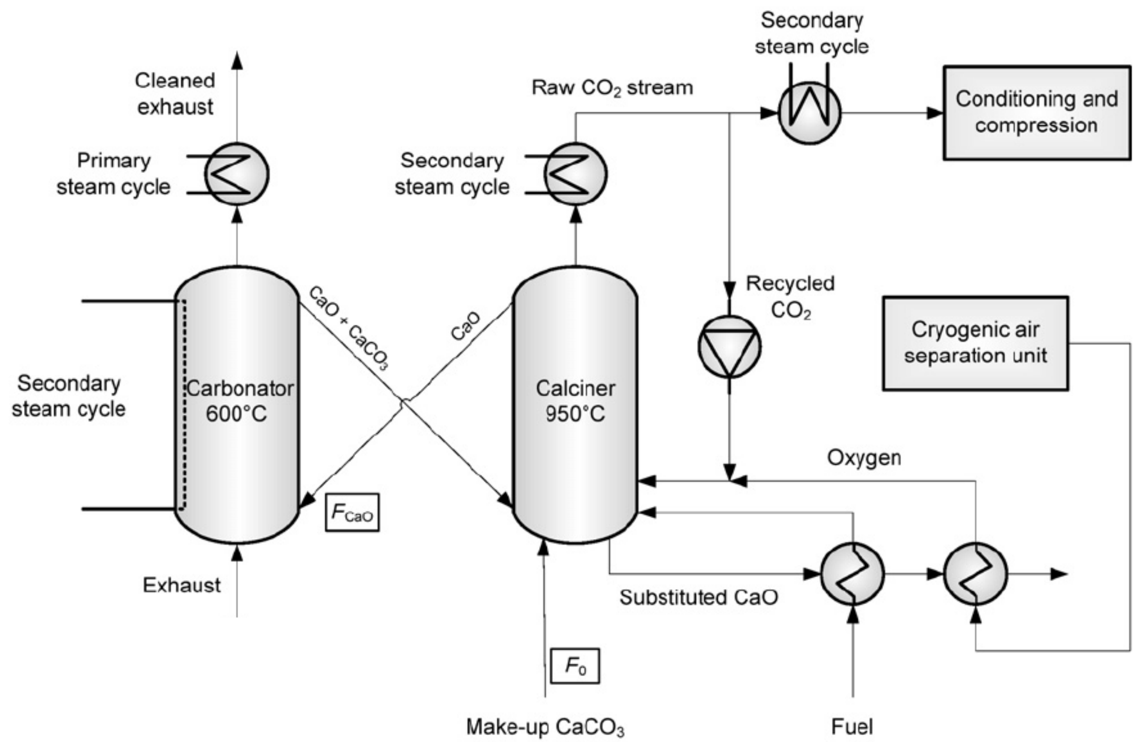
  - Ammonia
- Physical absorption solvents
  - Rectisol
  - Selexol
- Solid Looping
  - Calcium looping
- Cryogenic capture

# Calcium looping

- **Brief description**- flue gas is fed into a carbonator reactor operating at 600-700ºC and atmospheric pressure, where $CO_2$ reacts with CaO to be converted into $CaCO_3$. Solids from carbonator are sent to a second reactor (calciner operating at 900°C) where $CaCO_3$ is again decomposed into CaO and $CO_2$. $CO_2$ is then captured and the CaO is recirculated to the carbonator reactor. Assume for techno-economic modelling $CO_2$ output pressure of *ca*. 1 bar.
- **Technology status**- Technology has been piloted. Proposed in 1999 as a carbon capture specific process; optimisation and pilot-scale tests are currently underway. Significant research into developing calcium looping as a fundamental processs within cement manufacture.
- **Technology providers**- Alstom, CANMET Energy Technology Centre, CEMEX
- **Economic and market factors**- System is expected to be significantly cheaper than current methods. Cheap and abundant sorbent (limestone), harmless exhaust gas Low energy penalty and operational costs- considering it can generate steam from heat released in the carbonation reaction. FOAK premia for calcium looping applied to industrial CCS are not well understood.
- **Key barriers and challenges**- High decay in sorbent's capture capacity. Scale up of the technology needs to be addressed. High operating temperature and effective heat exchange designs.
- **Cost estimation –** MacKenzie *et al*. (2007). Economics of $CO_2$ capture using the calcium cycle with a pressurized fluidized bed combustor. Energy & Fuels 21 920–926, updated to account for inflation (HIS CERA PCCI ex. Nuclear), UK location (factor 1.2) and $1.6/£1 conversion.

# Calcium looping flow diagram

D. Berstad et al. / International Journal of Greenhouse Gas Control 11 (2012) 25–33



# References - Solid looping (1/2)


**Calcium looping**

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# References - Solid looping (2/2)


**Calcium looping**

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# $CO_2$ capture technologies

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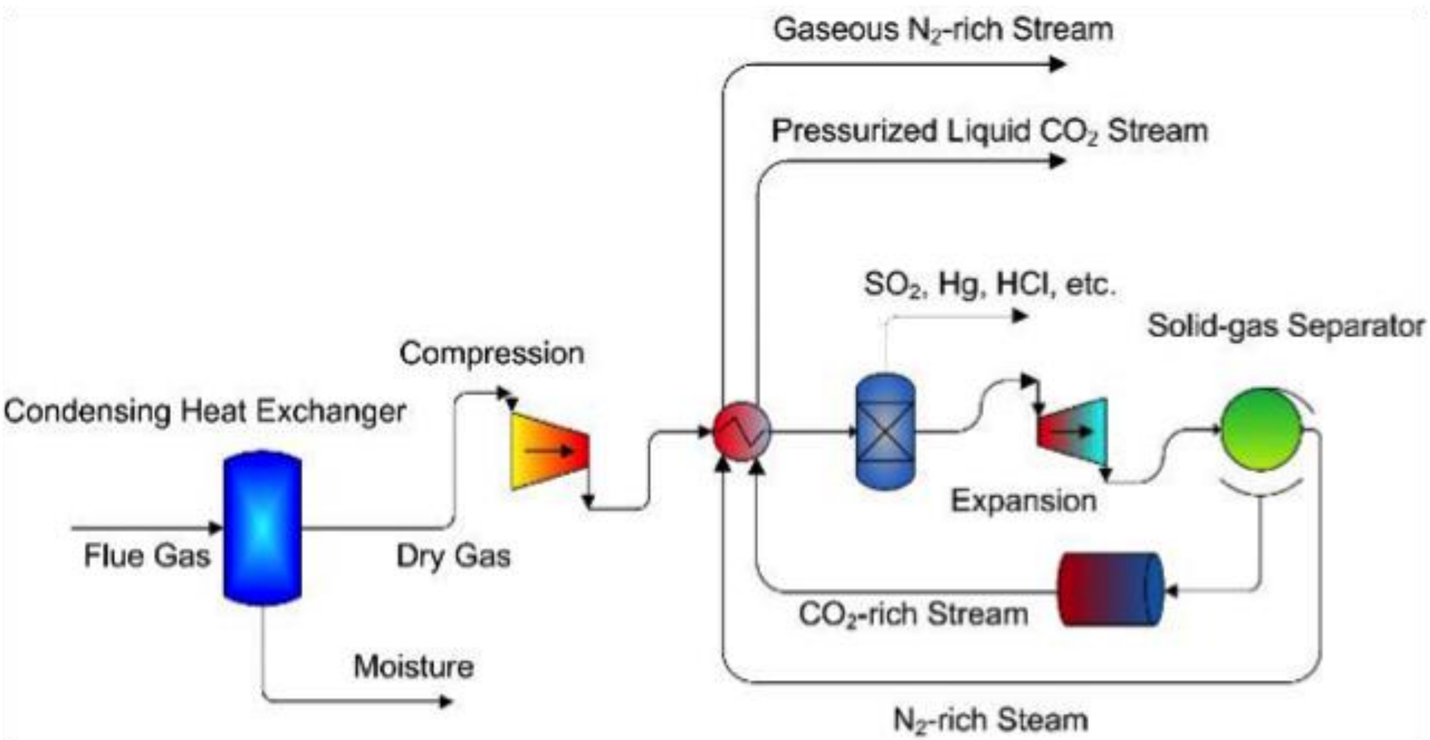
  - Calcium looping
- Cryogenic capture

# Cryogenics

- **Brief description**- utilizes the principle of separation based on cooling and condensation with gases that have different boiling temperatures. For $CO_2$ separation, the feed gas is cooled to a temperature below the sublimation temperature of $CO_2$ (c.a. -78°C) to desublimate $CO_2$ from the gas phase by forming solid $CO_2$ (dry ice).
- **Technology status**- used commercially for streams that already have high $CO_2$ concentrations (e.g. >90%), for instance in gas separation.
- **Technology providers**- Sustainable Energy Solutions; General cryogenic gas separation companies: Air products, Linde-BOC, Air Liquide, Cryogenmash, Cryotec Anlagenbau GmbH, Chart Inc, Costain
- **Economic and market factors**- produces a high gas purity output stream. Economically feasible for high concentration, high pressure gas streams. Only carbon capture method that does not require any $CO_2$ carrier material. It requires minimal changes to the existing plant, and offers the added value of removing NOx, $SO_2$, HCl, and Hg during the same process. Cost estimate based on Tuinier et al (2011), adjusted for inflation, UK conditions, and $1.64/£1 conversion. FOAK premia for cryogenic separation applied to industrial CCS are not well understood.
- **Key barriers and challenges**- very few studies cryogenic separation for dilute flue gas streams e.g. from post-combustion $CO_2$ capture. Substantial energy requirement makes it less desirable for applications with low partial pressure $CO_2$. Pre-filtering is required to avoid blockage when frozen (e.g. water). High operational cost and significant energy penalty.

# Conventional Cryogenic capture flow diagram for liquid $CO_2$




Step 1) Dry gas

Step 2) Cool gas through expansion

# Cryogenic capture diagram for solid or gaseous $CO_2$

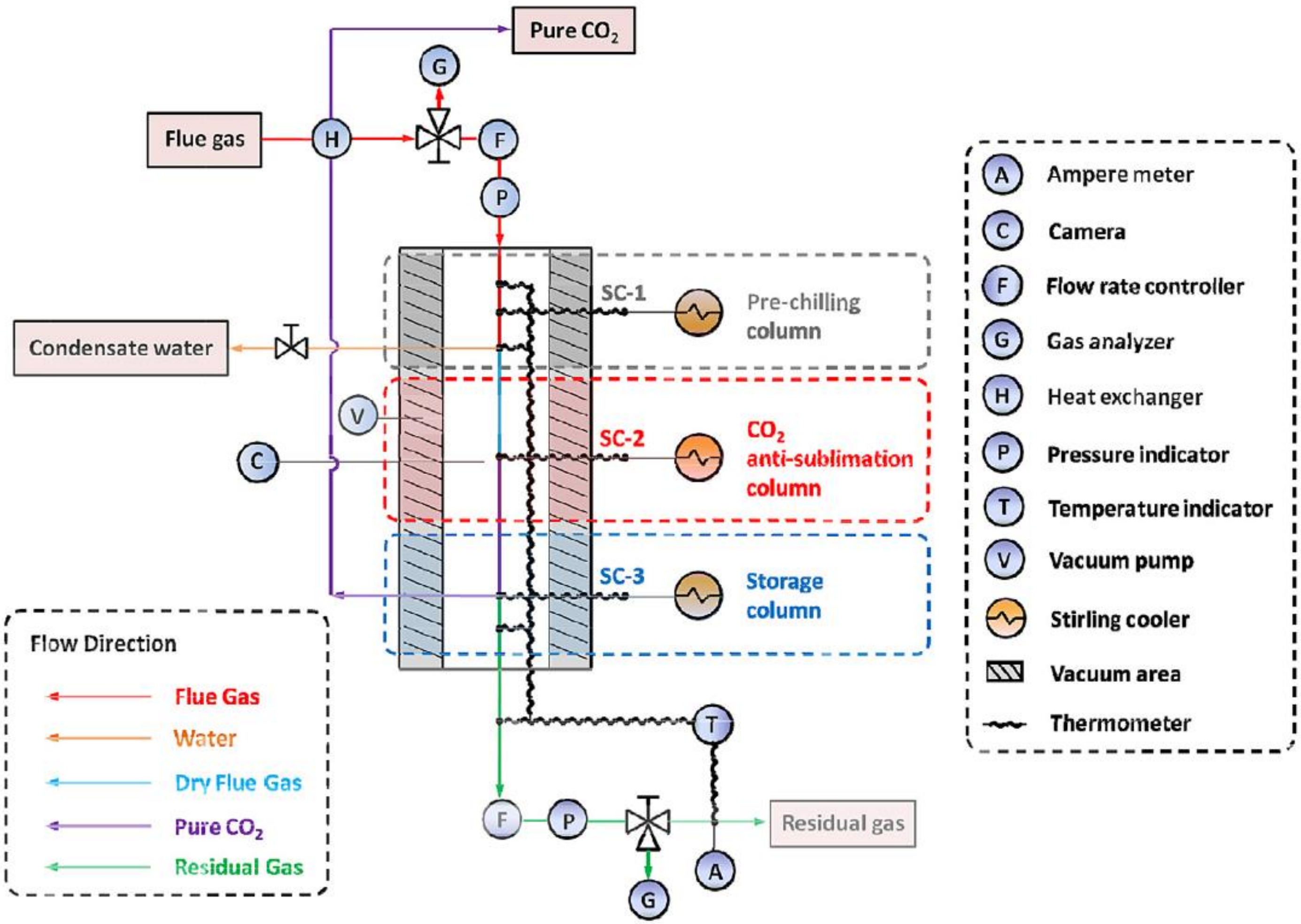



**Fig. 1.** Schematic of the developed anti-sublimation capture process.

# References - Cryogenic capture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# Outline

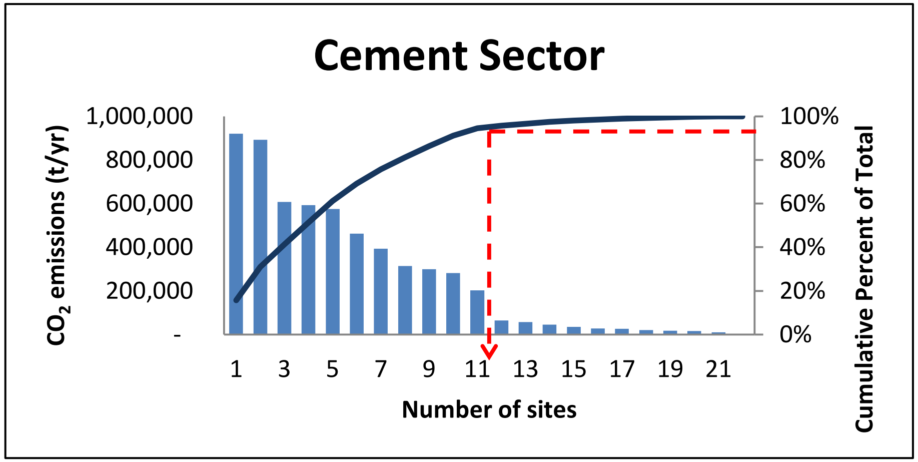
- Overall Project Methodology
- $CO_2$ capture technologies
- $CO_2$ sources
- Techno-economic analysis of industrial $CO_2$ capture
- Process simulation case studies
- $CO_2$ utilisation review

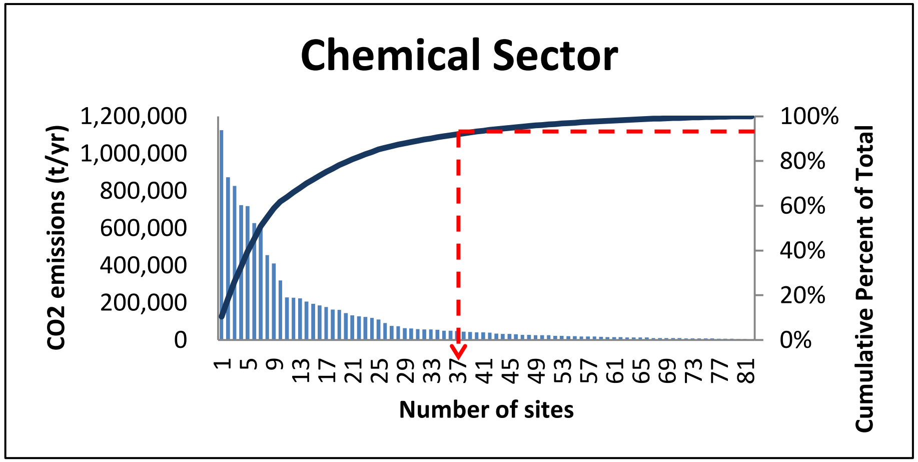
# $CO_2$ source databases are prepared from public and grey literature, and further refined with stakeholders and through process simulations

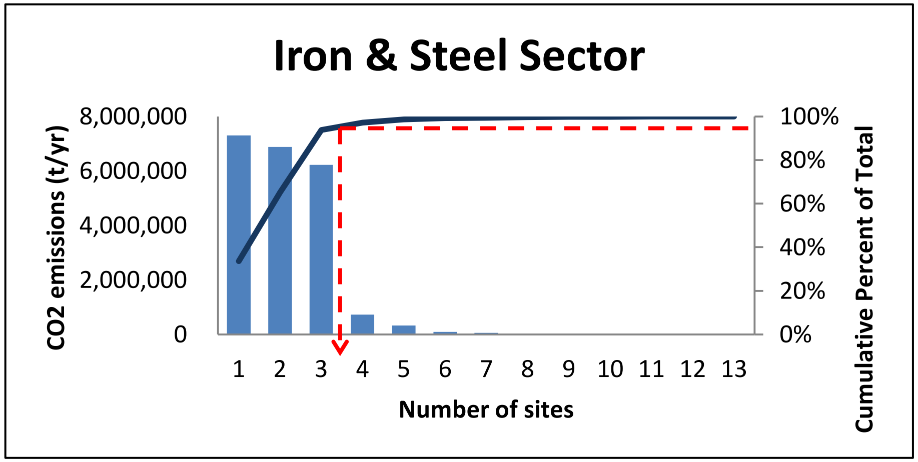
| Process | Key |
|---|---|
| Collate data sources | • **Latest $CO_2$ data sources are collected** (EU ETS, Environment Agency, SEPA, Element Energy in-house datasets, HSE, NI dataset, others) |
| Source long list | • Manual Quality Control through cross-referencing |
| Filtering of sources | • Identify most relevant sources in four sectors based on largest emissions (coverage of 80% of UK sectoral $CO_2$ emissions) |
| Source archetypes | • Techno-economic model design clarifies source data collection requirements<br>• Literature review to identify relevant source archetype properties |
| Populate sources database | • Focussed literature review **to populate database** |
| Review with stakeholders | • **Source database circulated** with industry experts for review of assumptions |
| Review with process simulations | • Source database and analysis results are **reviewed against bespoke process simulations** of capture processes in the different industries |

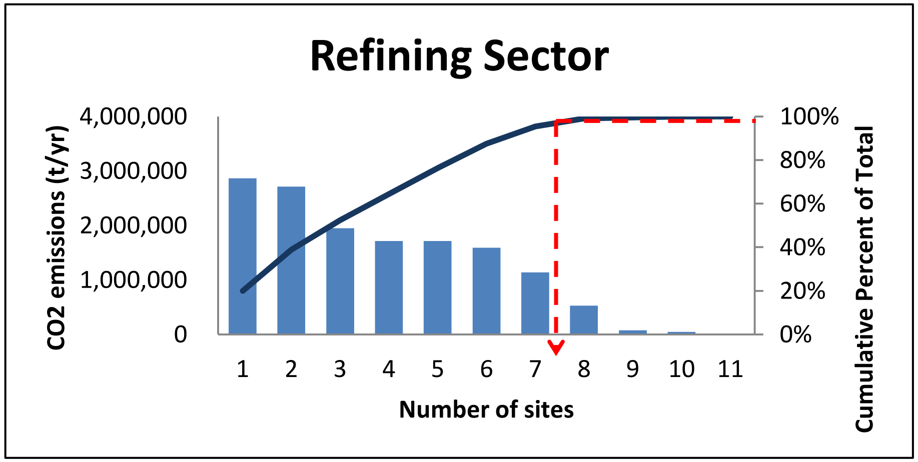
# 52 industrial $CO_2$ sources used in sources database.

Most relevant sources identified in four sectors based on largest emissions (coverage of 80% of UK sectoral $CO_2$ emissions).

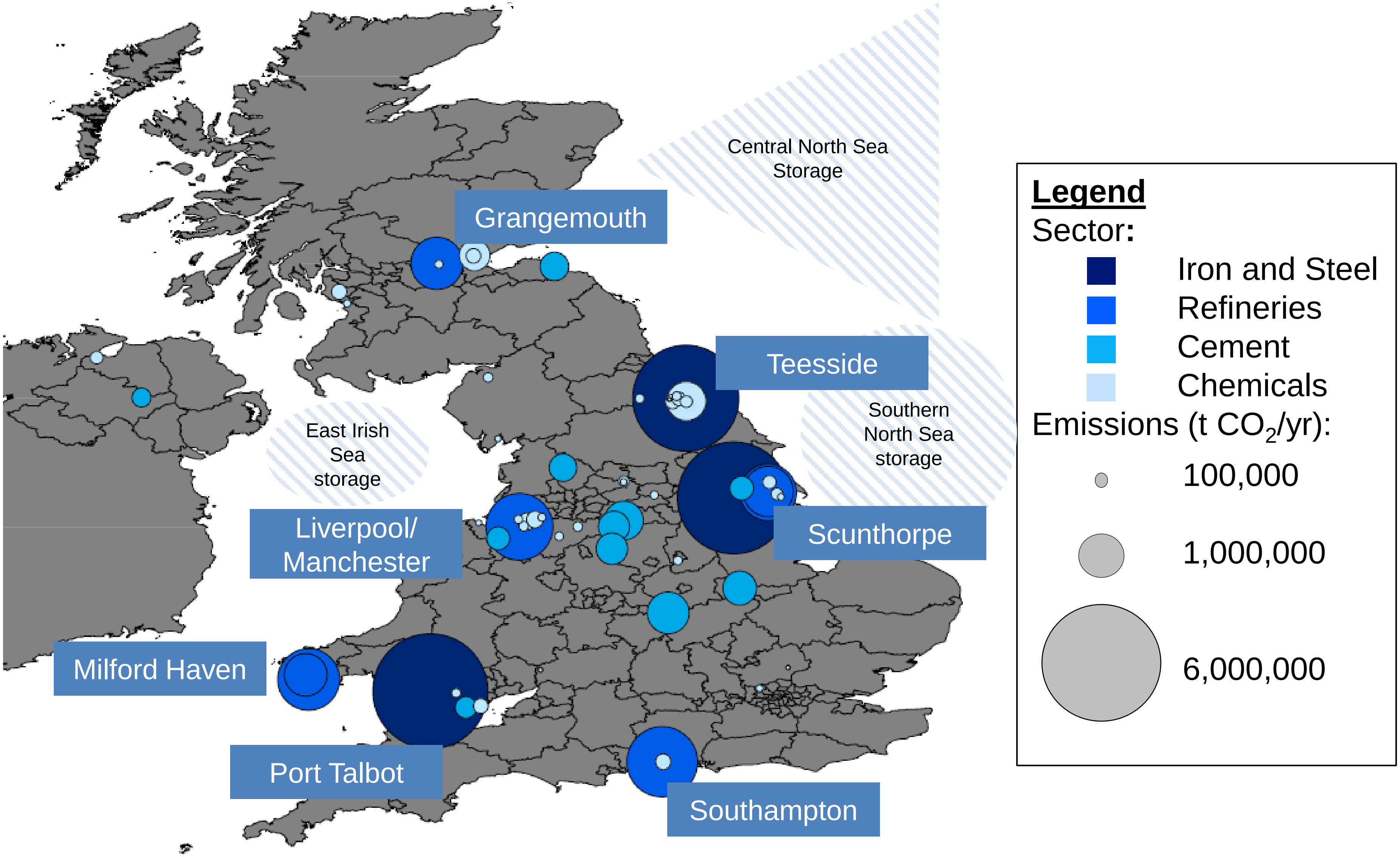


# 52 large industrial sites were included in the techno-economic modelling.



# Source properties identified in the sources database

1. Name
2. Location (lat/long)
3. Annual $CO_2$ emissions
4. Sector
5. Relative $CO_2$ emissions forecast for 2020 and 2025 (default is 100%)
6. Vent complexity (ideally number of vents)
7. Archetype properties for
   I. %$CO_2$
   II. Temperature
   III. Pressure
   IV. Indicative levels of key impurities currently vented ($NO_X$, $SO_X$, $N_2$, $O_2$, $CH_4$, CO, $H_2S$, $H_2O$, Particulate Matter (PM), other)
8. Site cooling water availability (low/high)
9. Site process water availability (low/high)
10. Site space available for capture plant (low/high)
11. Sub-sector cost complexity factor (multiplies capex and opex)
12. Traded or not-traded $CO_2$
13. Amount of waste heat available on site (Central/High)
14. Cost of heat available on site (Low/Central)
15. Familiarity with managing complex processes (Y/N)
16. Hidden cost e.g. downtime

# Source archetype properties

| Source archetype | Vent complexity | $CO_2$ stream pressure (MPa) | Input $CO_2$ (% volume) | | | Relative emissions in 2020 (2013 = 100%) | | | Relative emissions in 2025 (2013 = 100%) | | | Impurity level (ppm) | | %of source available |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | Low | Central | High | Low | Central | High | Low | Central | High | NOx | SOx | |
| Steel | Many | 0.11 | 16% | 30% | 44% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 60% |
| Cement | Single | 0.11 | 14% | 24% | 33% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 99% |
| Other Refinery | Many | 0.11 | 8% | 10% | 12% | 80% | 100% | 125% | 80% | 100% | 125% | 600 | 1200 | 90% |
| Coal power plant | Single | 0.11 | 8% | 11% | 15% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 99% |
| Gas power plant | Single | 0.11 | 2% | 3% | 6% | 80% | 100% | 125% | 80% | 100% | 125% | 10 | 10 | 99% |
| Hydrogen | Single | 0.11 | 20% | 95% | 99% | 80% | 100% | 125% | 80% | 100% | 125% | 10 | 10 | 99% |
| Gas boiler | Single | 0.11 | 5% | 7% | 10% | 80% | 100% | 125% | 80% | 100% | 125% | 10 | 10 | 99% |
| Industrial Coal CHP | Single | 0.11 | 8% | 12% | 15% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 99% |
| Industrial Gas CHP | Single | 0.11 | 2% | 3% | 6% | 80% | 100% | 125% | 80% | 100% | 125% | 10 | 10 | 99% |
| Ammonia - pure CO2 stream | Single | 5.1 | 90% | 95% | 100% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 99% |
| Crackers | Many | 0.11 | 8% | 10% | 12% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 99% |
| Other chemicals | Many | 0.11 | 2% | 11% | 40% | 80% | 100% | 125% | 80% | 100% | 125% | 100 | 100 | 99% |

# Indicative impurity inventory for $CO_2$ streams

| $CO_2$ source | Impurity assumptions |
|---|---|
| 1. For oil/coal using sites and majority of chemicals sites | • Up to maximum of 80% $N_2$, 20%$O_2$, 1% Ar, 20% water vapour.<br>• Depending on emissions control in place, up to low 10s to low 100s of ppm levels of NOx, SOx, $CH_4$, CO, PM, $H_2S$, VOC, and ppb levels of heavy metals. |
| 2. For gas combustion sources | • Up to maximum of 80% $N_2$, 20%$O_2$, 1% Ar, 20% water vapour.<br>• Up to 10s of ppm levels of NOx, SOx, $CH_4$, CO, PM, $H_2S$, VOC, CO, depending on emissions control in place.<br>• Variable water vapour. Ppb of heavy metals |
| 3. For $CO_2$ from steam methane reforming (hydrogen plants) | • Up to 10s of ppm levels of $N_2$, $O_2$, NOx, SOx, $CH_4$, CO, PM, $H_2S$, VOC, CO, $H_2$.<br>• Variable water vapour. Ppb of heavy metals |
| 4. For $CO_2$ from ammonia production | • Up to 100s of ppm levels of $N_2$, $O_2$, NOx, SOx.<br>• Up to 10s of ppm of $CH_4$, CO, PM, $H_2S$, VOC, CO, $NH_3$, depending on emissions control in place.<br>• Variable water vapour. Ppb of heavy metals |

# Cross-sectoral barriers to carbon capture technology adoption (1/3)

| Issue | Barriers |
|---|---|
| **High cost** | Carbon capture facilities require very high investment costs of hundreds of millions of pounds for the largest industrial sites, above typical site budgets. These costs are incurred in an environment of currently low and uncertain future revenues and investor risk aversion for non-core investments. First movers may be locked into high cost configurations. |
| **High cost uncertainty** | The uncertainties on the total costs for developing carbon capture plants in industry are very high. This results from the limited number of realised (industrial) carbon capture plants, the small scale of demonstrations compared to target commercial scale, as well as the early development stage of some of the technology components and site to site differences. |
| **Funding for scale up** | Limited to no funding is currently available for next phases of capture plant scale up (demos) in industrial sectors. |
| **Application not proven at scale** | Carbon capture is currently not proven at full scale for any industrial site. Further scale up can reveal additional project risks and complications. The main risks being additional project costs and operational impacts. |
| **Technologies not developed to commercial ready level** | Most of the capture technologies considered in this study are not yet developed to a commercially ready level (TRL9). The working of all of these technologies is proven beyond bench scale though, and across industry stakeholders few reservations were made inherently about the technologies themselves. |
| **Plant integration risks** | The process design in energy intensive industries is usually more complex than that of power plants, posing additional challenges in integrating capture plants with process facilities. The main plant integration challenges are;<br>Downtime (hidden costs). The integration of a capture plant in an existing process may require additional downtime of the facility, beyond regular overhaul periods. This can lead to additional costs, for instance due to missed revenues, additional maintenance facilities or the need to make other arrangements to ensure supply. The latter is especially relevant in the refining sector where a refinery sometimes supplies a specific area and alternative supply chains are not readily available.<br>For large continuously operated facilities the periods between major overhauls can be very long (around five years for fluid catalytic crackers in refineries and up to ten years for blast furnaces in the iron and steel sector). When a capture plant can only be reasonably brought online in a major overhaul this can represent limited windows of opportunity for the development of capture plants. |

# Cross-sectoral barriers to carbon capture technology adoption (2/3)

| Issue | Barriers |
|---|---|
| **Production/unavailability risks** | Extending an industrial process with a capture plant increases the complexity and operational dependencies of the overall facility. Across the different industries this increase in operational complexity is seen as a significant risk, especially for availability. Specific aspects include process transients, compatibility of different streams and erosion and corrosion issues. |
| **Impact on product quality** | A capture plant can have an adverse impact on product quality. Especially when it impacts the process conditions and operation of the main plant. |
| **Unfamiliarity with CCS technologies** | The main barriers to the deployment of carbon capture technologies within industrial sectors are the high costs and the limited commercial need to develop these applications. As a result of this the energy intensive industry has little experience and limited familiarity with carbon capture technologies, especially compared to the power sector. Different sectors have different levels of familiarity and experience with specific types of processes (gas separation, solids handling) employed in CCS technologies. This can potentially reduce or increase this barrier for specific technology-sector combinations, which are addressed in the sector specific barriers. |
| **Data sharing / knowledge gaps** | As few carbon capture plants have been developed in industry to date, there is a lack of company and site expertise which is exacerbated by issues around data sharing between companies. |
| **Large differences between sites limit replicability of solutions** | Especially for the chemicals, oil refining and iron and steel sectors the actual layout and process design of different facilities within one (sub) sector can vary strongly, limiting knowledge transfer and replicability of solutions across sites. |
| **Limited sector specific process understanding** | There are only limited detailed process simulations for application in specific processes, as well as limited trials in specific processes. |
| **Effects of impurities** | Similar to the above barrier on specific process understanding, the effects of different impurity conditions across different plant types is not yet fully understood and investigated. The main impurities of concern are SO2 (amines for instance react with acidic compounds and form amine salts that don’t dissociate in the stripper), NOx (solvent degradation) and particulates (affects for instance an amine CO2 absorber). |

# Cross-sectoral barriers to carbon capture technology adoption (3/3)

| Issue | Barriers |
|---|---|
| **Uncertainty on long term availability of facility increases risk of capture investment** | Many industrial facilities have already operational for several decades and there is an inherent uncertainty of the remaining lifetime of any one site. Especially for facilities which provide little margin or run at a loss, this is a key uncertainty, and under current conditions including a post-combustion carbon capture facility may reduce margins even further. Examples of this have been Longannet coal power plant in the UK and the Florange steel plant in France (ULCOS). At both these sites, capture plants were being developed for sites with challenging fundamental economics.<br>This uncertainty is even stronger in the energy intensive industries, which supply into global competitive markets. There is always uncertainty whether an industry and specific facilities will still be profitable and operational in the current locations in the future. This uncertainty is less present in the power sector, as electricity production is required more locally. |

# Source sectors

- Cement
- Chemicals
- Iron and Steel
- Oil Refining
- Other sectors of potential relevance

# Cement – Sector description

## Sector context

The UK cement sector consists of 6 companies with a total of 13 sites. Of these, 11 cement plants are producing clinker as of 2012. Typical UK cement production is 10 million tpa.

## Description of main processes

Cement production can be divided in two basic steps. (1) Clinker is made in a rotary kiln at temperatures of 1450°C, after which (2) Clinker is ground with other minerals to produce the powder we know as cement. Raw materials are limestone (for lime), clay, marl or shale (for silica, alumina, and ferric oxide) and other supplementary materials such as sand, pulverised fuel ash (PFA), or ironstone (to achieve the desired bulk composition). More and more low and zero carbon waste fuels are being used. Most of the plants in the UK are of the dry process type (grinding mineral components without addition of water).

## UK sites

The UK had 11 cement plants producing clinker in 2012 with emissions ranging from 0.2 to 1.1 $MtCO_2/yr$.

# Cement in the UK– $CO_2$ sources, sites and barriers

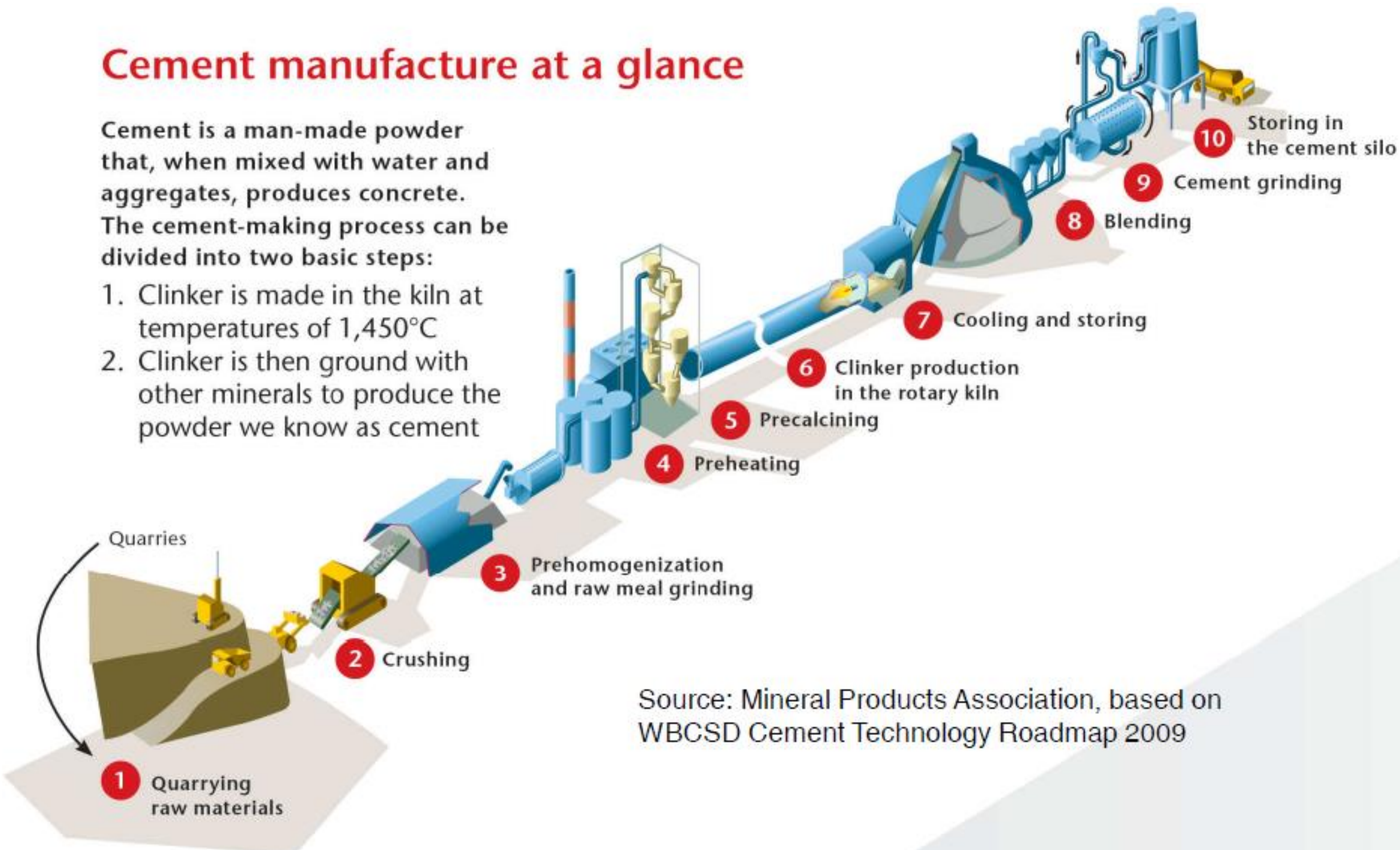
## $CO_2$ sources and emissions

Cement production has two major $CO_2$ emission points: fuel combustion to provide process heat (30%) and process related emissions (60%; due to the decomposition of the $CaCO_3$).

Since 1990, an absolute reduction of 54% in emissions has been achieved, based on efficiency, fuel switching and changes in output.

| CO2 emission point | Description | % of total emissions | % concentration of CO2 stream | Notes |
|---|---|---|---|---|
| Limestone calcination | Process used to convert limestone to lime, one of the key components of cement. | 60% | 24% | Calciner flue gas at ~850°C |
| Heat/power | Heat is required for calcination and to run the cement kiln. | 30% | 2% - 15% | % of $CO_2$ concentration depends on fuel |

# Cement – illustrative rotary kiln-based cement manufacture




Source: Mineral Products Association, based on WBCSD Cement Technology Roadmap 2009

# Case study of a CCS initiative in the Cement Sector

- The IEA CCS Cement Sector CCS Roadmap (2009) identifies that multiple CCS demonstration projects are required worldwide by the early 2020s to build capacity and allow the sector to achieve deep cuts in emissions by 2050.
- The European Cement Research Academy (ECRA) is supporting a five phase approach to CCS development. Reports from the first two phases identify post-combustion, oxyfuel, membranes and calcium carbonate looping as technologies of interest for cement CCS.
- The ECRA Phase II report describes cost estimates, infrastructure requirements, and technology challenges for post-combustion capture (e.g. using MEA) and oxyfuel capture based on high level engineering studies, and compared with work by MottMacDonald for the IEA Greenhouse Gas R&D Programme (2008).
- Results from the third phase (lab/small-scale research activities) are expected to be published later this year. The next phase involves pilot scale research activities (ca. £10m), including post-combustion and oxyfuel for around two years at the Norcem Brevik site in Norway. Precise details of the scale and configurations involved are not yet in the public domain. The piloting activity is expected to be followed by demonstration plant scale projects are expected towards the end of the 2010s.

# Cement - technology barriers for deploying a capture project in the period 2020 to 2025

| Technology | Barriers | |
|---|---|---|
| **General** | • Very few worldwide piloting or demonstration projects underway or planned.<br>• Some sites are in locations where new industrial development, access to $CO_2$ transport networks, or cooling water availability is highly restricted.<br>• Risk of technology and process lock-in to a high cost solution.<br>• Fuel switching and volatile output quantities depending on economic cycle make capacity management difficult. | |
| **1st generation amine solvents** | • Few studies of compatibility of flue gas with capture stream and the least cost pretreatment solutions required.<br>• High capex and opex (incl. heat demand) in excess of typical site expenditures.<br>• Logistics and HSE challenges associated with amine storage and manipulation, likely to elevate COMAH status.<br>• Limited cooling water availability could restrict potential at some sites. | • Lack of familiarity with solvent-based gas separation technologies.<br>• Challenge to synchronise integration with plant downtime.<br>• Contaminants in flue gas results in degradation (NOx) and salt formation (SOx), resulting in a high amine requirement |
| **2nd generation chemical solvents (e.g. advanced amines, amino acids and blends)** | • Focus of capture technology development is for power sector – not clear to what extent solvent development is targeted at improving compatibility with cement production.<br>• Similar to first gen amine solvent technologies, though expect costs, footprint, water, heat, and HSE impacts expected to be less severe. | • Multiple chemicals are under development at TRL6 , so concepts need to be proven through pilots and demos and there is a risk of cost and performance issues arising. |
| **Chilled ammonia** | • Usually no ammonia on site, would introduce new risks | |
| **Chemical looping (e.g. calcium looping)** | • TRL6 technology requiring significant piloting and demonstration before it can be applied at scale of 100,000t/yr or higher. Few suppliers. Available performance models need refining. | • Extent of calcium looping integration with core process is unclear – ideally would source hot $CO_2$ rich flue gas at high temperature directly from kiln rather than at the end of the process, but this would involve overhaul of site. Not clear if by-product salts can be sold. |
| **Oxyfuel capture** | • Good long-term cost reduction potential but need for baseline process redesign, with multiple site impacts including potentially kiln management, flow management, change of fuel supply, core product specification. Sealing methods not well demonstrated. Extensive system and component piloting required to manage project-on-project risks. | • However, there are expectations that these barriers can eventually be solved, although not necessarily in time for operation in 2025.<br>• The higher flame temperature with oxyfuel combustion may lead to reduced overall system costs. |

# Cement Sector $CO_2$ Source Database Assumptions

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2$/yr) | COMAH |
|---|---|---|---|---|---|---|---|
| Cement | Cemex UK Cement Limited | Rugby Works | Cement | 52.38 | - 1.29 | 1,065,000 | None |
| Cement | Hope Cement Ltd | Hope Cement Works | Cement | 53.34 | - 1.75 | 922,000 | None |
| Cement | Lafarge Tarmac Trading Limited | Lafarge Tarmac Tunstead Cement | Cement | 53.28 | - 1.86 | 578,000 | None |
| Cement | Castle Cement Ltd (Hanson) | Ketton Works | Cement | 52.64 | - 0.55 | 698,000 | None |
| Cement | Lafarge Tarmac Cement and Lime Ltd | Lafarge Tarmac Cauldon | Cement | 53.04 | - 1.87 | 597,000 | None |
| Cement | Lafarge Tarmac Cement and Lime Ltd | Lafarge Tarmac Dunbar | Cement | 55.98 | - 2.47 | 491,000 | None |
| Cement | Castle Cement Ltd (Hanson) | Padeswood Works | Cement | 53.16 | - 3.06 | 310,000 | None |
| Cement | Castle Cement Ltd (Hanson) | Ribblesdale Works | Cement | 53.89 | - 2.39 | 441,000 | None |
| Cement | Cemex UK Cement Limited | South Ferriby Works | Cement | 53.68 | - 0.53 | 334,000 | None |
| Cement | Lafarge Tarmac Cement and Lime Ltd | Lafarge Tarmac Aberthaw | Cement | 51.39 | - 3.39 | 286,000 | None |
| Cement | Lafarge Tarmac Cement and Lime Ltd | Lafarge Tarmac Cookstown | Cement | 54.63 | - 6.73 | 231,000 | None |

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# Source sectors

- Cement
- Chemicals
- Iron and Steel
- Oil Refining
- Other sectors of potential relevance

# Chemicals - Sector description

## Sector context

The chemicals industry comprises many different products, ranging from (in)organic bulk chemicals to small volume special products. A number of processes in industry result in a high purity and high concentration $CO_2$ exhaust gas, which can be readily captured . These processes include hydrogen production and some organic chemical production processes (e.g. ethylene oxide production). Hydrogen production processes are also used in ammonia (and ammonia based fertiliser) production, and methanol. On a global scale, the $CO_2$ emissions from these activities are relatively low compared to emissions from other industrial activities, but these $CO_2$ streams offer an early-adoption opportunity for CCS demonstration projects.

## Description of main processes

Due to the wide ranges of different chemicals produced, various combinations of processes take place in the chemical sector. The description below focuses on high concentration CO2 sources.

Globally, around 45 - 50 million tonnes (Mt) of hydrogen are produced each year, the majority of which is produced from fossil fuels. Around half is used to produce ammonia and around a quarter is used for hydrocracking in petroleum refining.

Ammonia is typically produced using the Haber Bosch process, which starts with hydrogen production. The hydrogen is then reacted with nitrogen from air to form ammonia, producing a near-pure stream of CO2. (Around 80% of all ammonia manufactured worldwide is used to produce inorganic nitrogen based fertilisers.)

## UK sites

Over 30 chemical sites in the UK with varied flue gas mixtures.

High purity CO2 sources:

– 3 ammonia sites

– 1 hydrogen site

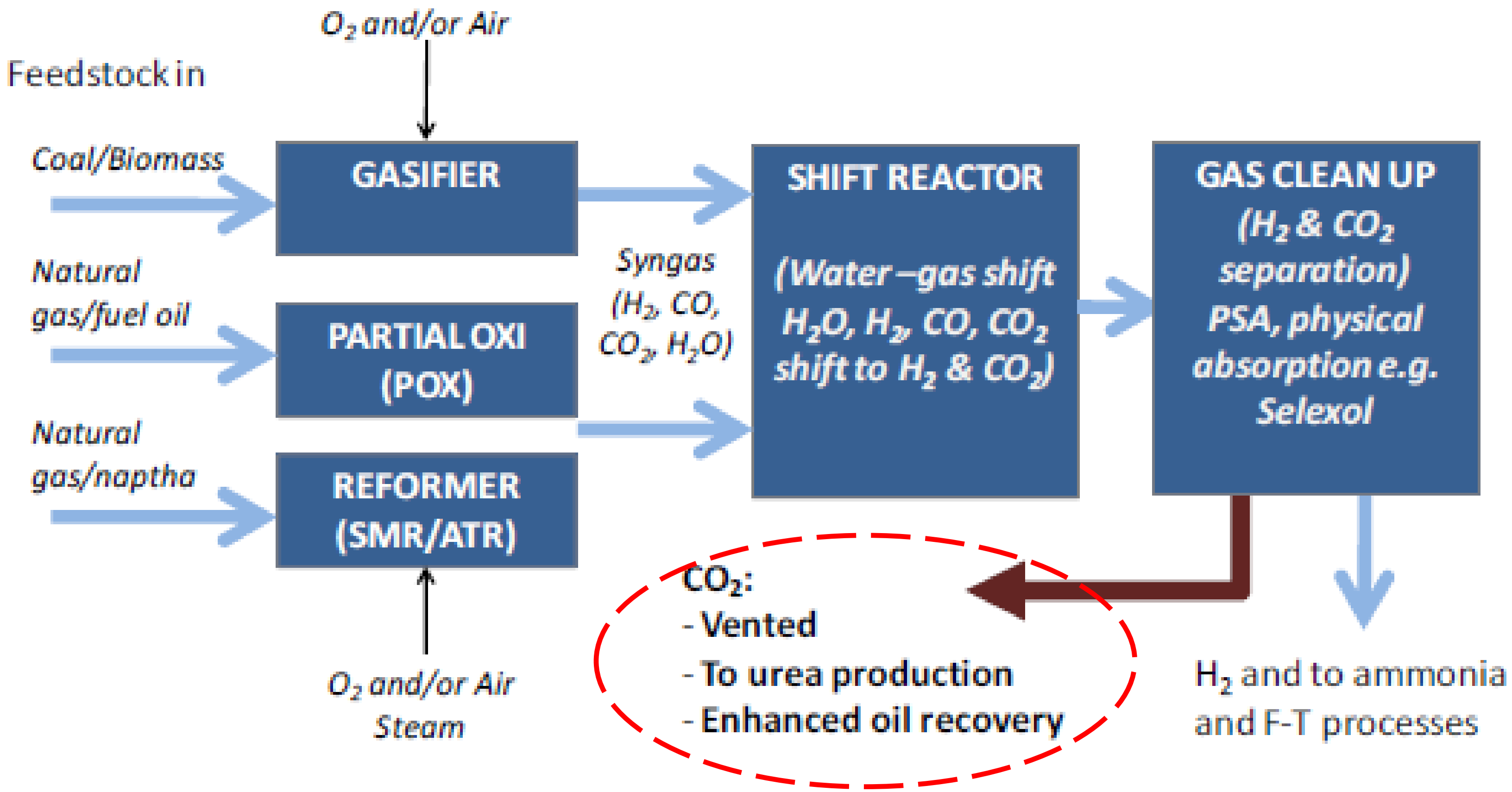
# Chemicals in the UK– $CO_2$ sources, sites and barriers

## $CO_2$ sources and emissions

High purity sources (hydrogen and ammonia)

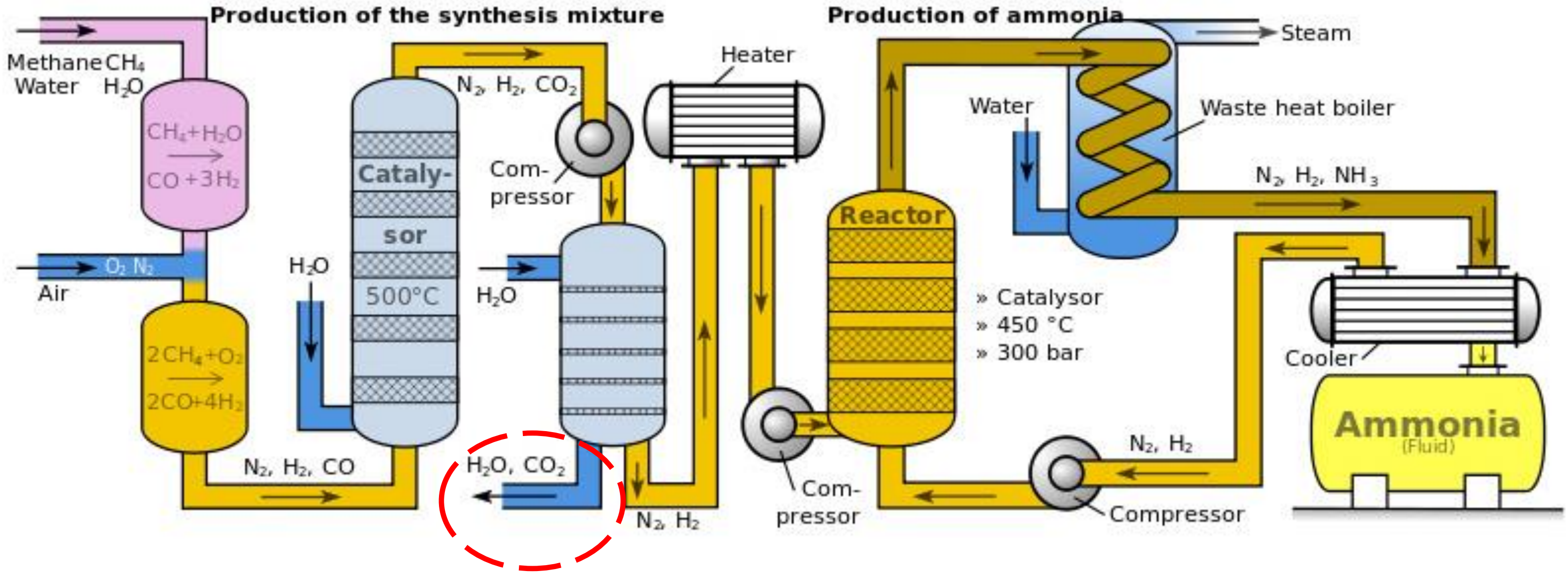
| CO2 emission point | Description | % concentration of $CO_2$ stream | Notes |
|---|---|---|---|
| Hydrogen | Hydrogen can be produced via gasification, partial oxidation or steam reforming. | 20-99% | |
| Ammonia | Hydrogen production is the first step in manufacturing ammonia | 90-100% | High pressure CO2 |
| Crackers | | 8-12% | |
| Other chemicals | | 8%-40% | |

# Hydrogen production flow diagram




Source: United Nations Industrial Development Organization (UNIDO). (2010). *Carbon Capture Applications : and Storage in Industrial Applications: Technology Synthesis Report Working Paper - November 2010*. p14

# Ammonia production flow diagram



# Chemicals - technology barriers for deploying a capture project in the period 2020 to 2025

| Technology | Barriers | |
|---|---|---|
| **1st generation amine solvents** | • Diverse industrial $CO_2$ streams, with each product exposed to different market forces.<br>• Some sites will have many furnaces which may be difficult to collect from.<br>• Gas powered furnaces have dilute streams.<br>• Baseline of fuel efficiency, fuel switching and process redesign to minimise $CO_2$ production complicate planning for CCS.<br>• Very few studies of compatibility of flue gas with capture streams from chemical industry with amine capture and implications for the least cost pretreatment solutions required.<br>• High capex and opex (incl. heat demand) in excess of typical site expenditures.<br>• Limited cooling water restricts potential at some sites. | • Logistics and HSE challenges associated with amine storage and manipulation, likely to elevate COMAH status for some sites.<br>• Some sites are in locations where new industrial development, access to $CO_2$ transport networks, or cooling water availability is highly restricted.<br>• Very little worldwide piloting or demonstration scale activity underway or planned.<br>• Risk of technology and process lock-in to a high system cost solution, compared to alternative process<br>• Contaminants in flue gas results in degradation (NOx) and salt formation (SOx), resulting in a high amine requirement. |
| **2nd generation chemical solvents (e.g. advanced amines, amino acids and blends)** | • Focus of technology development is for power sector – not clear to what extent solvent development is targeted at improving compatibility with chemical production.<br>• Similar barriers to first gen amine solvent technologies, though expect costs, footprint, water, heat, and HSE impacts expected to be less severe. | • Multiple chemicals are under development at TRL6 , so concepts need to be proven through pilots and demos and there is a risk of cost and performance issues arising. |
| **Chemical looping (e.g. calcium looping)** | • TRL6 technology requiring significant piloting and demonstration before it can be applied at scale of 100,000t/yr or higher.<br>• Appears to be no industrial or academic interest in integrating with chemical furnace sites.<br>• Typically sites won't have much experience with calcium looping technologies.<br>• Few suppliers. | • Available performance models need refining.<br>• Extent of calcium looping integration with core process is unclear – ideally would source hot CO2 rich flue gas at high temperature directly from furnace rather than after cooling.<br>• Not clear if by-product salts can be sold.<br>• Less experience with handling solids, compared to fluid and gases |
| **Cryogenics** | • Requires a source of cooling to be competitive | |
| **Oxyfuel capture** | • Need for baseline process redesign, with multiple site impacts including potentially furnace management, flow management, change of fuel supply, core product specification. Sealing methods not well demonstrated. Creates project-on-project risk. | • However, there are expectations that these barriers can eventually be solved, although not necessarily in time for operation in 2025.<br>• The higher flame temperature with oxyfuel combustion may lead to reduced overall system costs. |

# Chemicals - sector specific barriers

| Issue | Barriers | |
|---|---|---|
| **High purity streams (hydrogen and ammonia production)** | • Multiple streams with high and low different purities. (Low purity streams are gas combustion) | • Key challenge is transport to store and ensuring compatibility with transport specification. |
| **Multiple source exhaust** | • Many distributed emission points across a site.<br>• Either large duct network at significant investment and blower operational cost as well as integration challenges or high investment in multiple capture plants. | • Units like crackers are not single sources, many vents; need to be brought together |
| **Lack of data** | • Very little public domain data on CO2 emissions at individual vents<br>• Lack of shared understanding of how CCS compares with alternative options for sites and the sector as a whole. | |

# Chemicals Sector $CO_2$ Source Database Assumptions (1/3)

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2$/yr) | COMAH |
|---|---|---|---|---|---|---|---|
| Chemicals | Wilton Olefins 6 (Cracker) | Sabic UK Petrochemicals Limited - 1 | Cracker | 54.58 | -1.1 | 1,030,000 | Top |
| Chemicals | SembCorp Utlilities Teesside Power Station | SembCorp Utilities Teesside Power Station | Industrial Gas CHP | 54.59 | -1.12 | 873,000 | Top |
| Chemicals | Winnington CHP (Brunner Mond) | Winnington CHP (Brunner Mond) | Industrial Gas CHP | 53.27 | -2.53 | 719,000 | None |
| Chemicals | Fife Ethylene Plant | ExxonMobil Chemical Limited | Cracker | 56.1 | -3.3 | 645,000 | Top |
| Chemicals | Fawley cogen | Fawley cogen | Industrial Gas CHP | 50.83 | -1.35 | 627,396 | None |
| Chemicals | Billingham, GroHow Ltd | Billingham, GroHow Ltd - 1 | Ammonia - pure CO2 stream | 54.58 | -1.27 | 455,000 | None |
| Chemicals | Kemira GrowHow UK Ltd. | Ince - 1 | Ammonia - pure CO2 stream | 53.28 | -2.8 | 411,000 | Top |
| Chemicals | Yara fertiliser plant operated by BP | Made at BP Saltend, Hull | Ammonia - pure CO2 stream | 53.74 | -0.24 | 320,000 | Top |
| Chemicals | Lucite International Billingham | Lucite International Specialty Polymers & Resins Ltd | Other chemicals | 54.59 | -1.28 | 228,050 | Top |
| Chemicals | Billingham, GroHow Ltd | Billingham, GroHow Ltd - 2 | Industrial Gas CHP | 54.58 | -1.27 | 228,000 | Top |
| Chemicals | North Tees, BOC Group Plc | North Tees, BOC Group Plc | Hydrogen via SMR | 54.6 | -1.19 | 223,000 | None |
| Chemicals | Kemira GrowHow UK Ltd. | Ince - 2 | Industrial Gas CHP | 53.28 | -2.8 | 206,000 | Top |
| Chemicals | Runcorn Halochemicals Manufacturing | INEOS ChlorVinyls Limited | Other chemicals | 53.35 | -2.68 | 194,000 | Top |
| Chemicals | NPOWER COGEN (HYTHE) LIMITED1 | NPOWER COGEN (HYTHE) LIMITED | Industrial Gas CHP | 50.83 | -1.34 | 170,000 | Top |

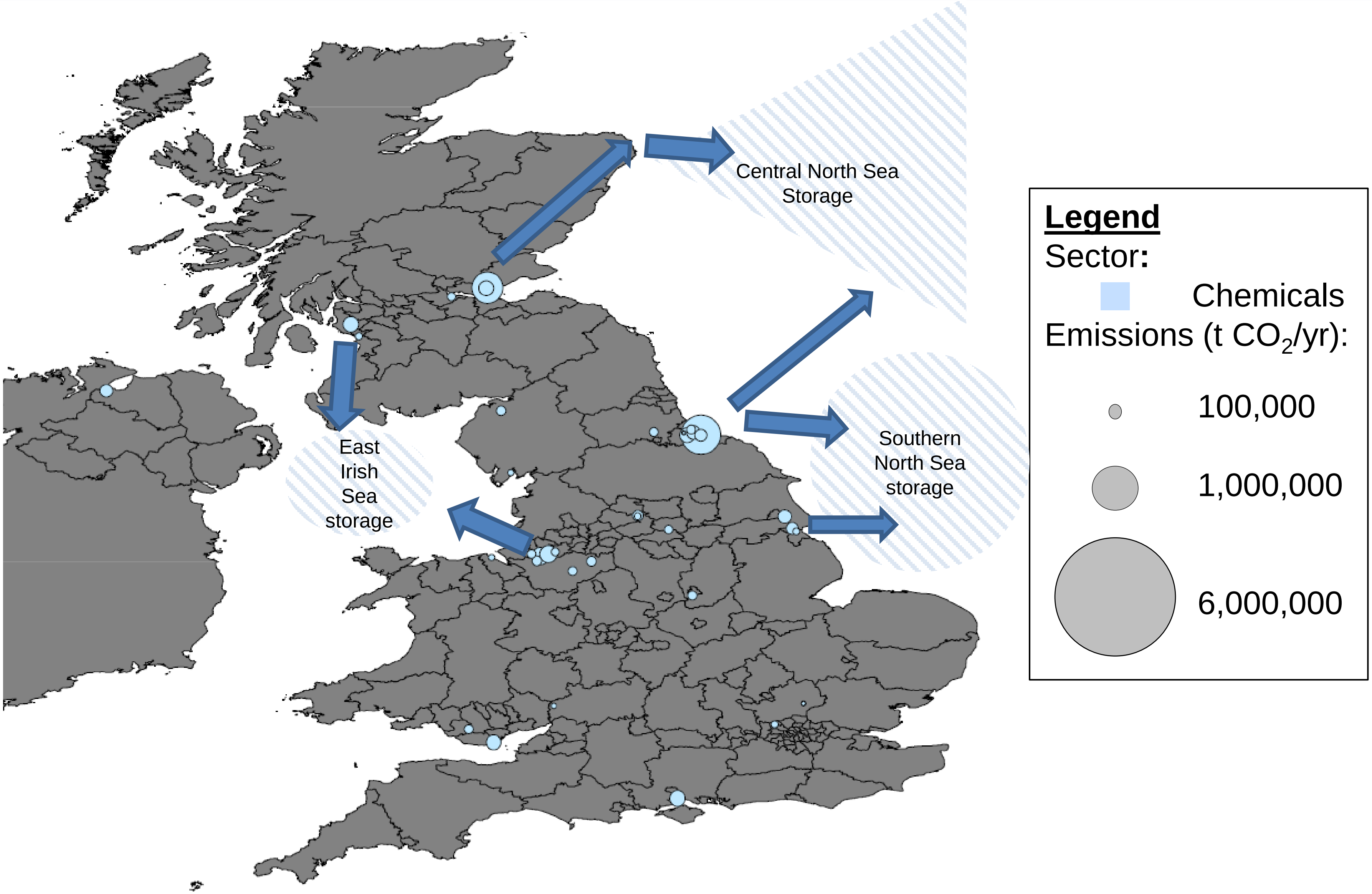
# Chemicals Sector $CO_2$ Source Database Assumptions (2/3)

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2/yr$) | COMAH |
|---|---|---|---|---|---|---|---|
| Chemicals | DSM Dalry1 | DSM Dalry | Other chemicals | 55.72 | -4.71 | 163,000 | Lower |
| Chemicals | Dow Corning Cogen Plant | Npower Cogen Limited - 1 | Industrial Gas CHP | 51.41 | - 3.24 | 158,000 | Lower |
| Chemicals | INEOS CHP PLANT | Npower Cogen Limited - 2 | Industrial Gas CHP | 54.58 | - 1.25 | 157,000 | Lower |
| Chemicals | Shell UK Ltd Fife NGL Plant1 | Shell UK Ltd Fife NGL Plant | Other chemicals | 56.09 | - 3.31 | 151,000 | None |
| Chemicals | Lotte Chemicals UK | Lotte Chemicals UK | Other chemicals | 54.58 | - 1.10 | 143,915 | None |
| Chemicals | BASF, Seal Sands | Ineos Nitriles (U.K) Ltd | Other chemicals | 54.61 | - 1.18 | 121,000 | Top |
| Chemicals | BP Chemicals Ltd, Hull | BP Chemicals Ltd | Other chemicals | 53.74 | - 0.24 | 121,000 | None |
| Chemicals | Invista UK Power Facility | Invista Textiles (UK) Ltd | Industrial Gas CHP | 55.03 | - 7.24 | 106,000 | Top |
| Chemicals | Millennium Inorganic Chemicals Ltd | Cristal Pigmanet Ltd | Other chemicals | 53.61 | - 0.16 | 100,000 | Top |
| Chemicals | North tees Aromatics | Sabic UK Petrochemicals Limited - 2 | Other chemicals | 54.58 | - 1.10 | 96,000 | Top |
| Chemicals | BASF Performance Products Plc - Bradford | BASF Performance Products plc | Other chemicals | 53.75 | - 1.76 | 79,000 | Top |
| Chemicals | Macclesfield | AstraZeneca UK Limited | Other chemicals | 53.28 | - 2.23 | 71,000 | Top |
| Chemicals | Kemira GrowHow UK Ltd. | GrowHow UK Limited, Ince Gas combustion | Gas boiler condensing | 53.28 | - 2.79 | 65,000 | Top |
| Chemicals | Wigton Boiler Plant | Innovia Films Ltd | Gas boiler condensing | 54.83 | - 3.16 | 65,000 | Top |
| Chemicals | Rockwool Bridgend | Rockwool Limited | Other chemicals | 51.55 | - 3.50 | 62,000 | Top |

# Chemicals Sector $CO_2$ Source Database Assumptions (3/3)

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2$/yr) | COMAH |
|---|---|---|---|---|---|---|---|
| Chemicals | British Salt Ltd. Middlewich Site | British Salt Limited | Other chemicals | 53.18 | - 2.43 | 61,000 | Top |
| Chemicals | Hydro Polymers Ltd | Ineos Newton Ayecliffe LTD | Other chemicals | 54.61 | - 1.59 | 59,000 | Top |
| Chemicals | D200 Energy Centre | Alliance Boots Holdings Limited | Industrial Gas CHP | 52.92 | - 1.19 | 59,000 | Top |
| Chemicals | Tioxide Europe Limited | Tioxide Europe Limited | Other chemicals | 54.63 | - 1.20 | 59,000 | Top |
| Chemicals | Polimeri Europa UK Ltd | Polimeri Europa UK Ltd | Other chemicals | 56.00 | - 3.68 | 50,000 | Top |

# Transport and storage opportunities most likely to be available in 2025 for chemical sector sources in Scotland, Teesside, Yorkshire and NW England.



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# Source sectors

- Cement
- Chemicals
- Iron and Steel
- Oil Refining
- Other sectors of potential relevance

# Iron and Steel – Sector description

## Sector context

The UK iron and steel sector can be subdivided into Integrated sites (only 3 sites – Teesside Scunthorpe and Port Talbot), and others (Electric Arc Furnace-only sites, mini-mills and several small re-rollers and annealers). In terms of carbon capture significance, focus is on Integrated Steelworks. In 2012, about 7.5 million tonnes of crude steel was produced in Integrated Steelworks. 2012 was atypical as Tata was rebuilding the blast furnace at Port Talbot and SSI were ramping up production in Teesside.

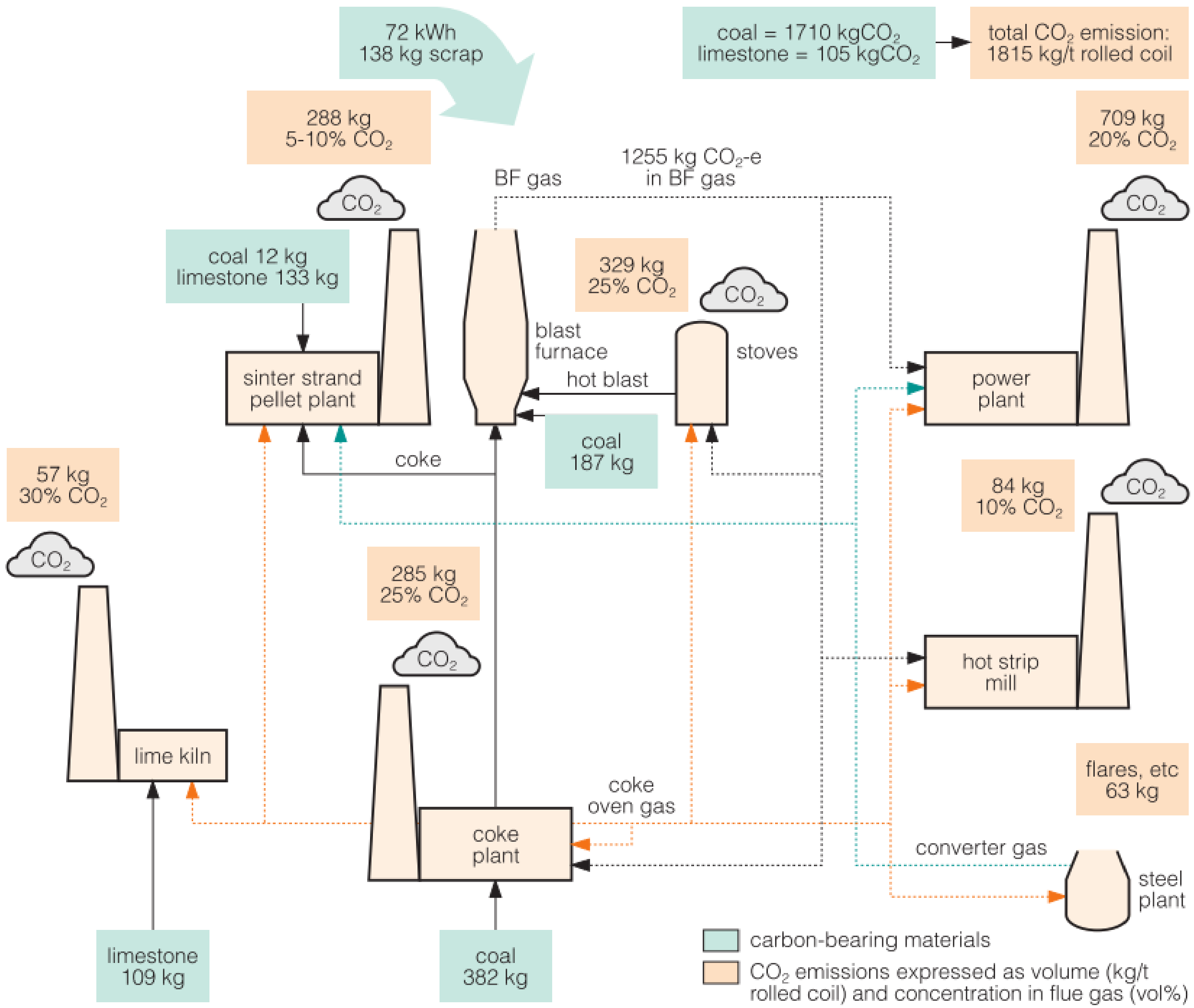
## Description of main processes

There are two process routes for making steel in the UK today: through an Electric Arc Furnace (EAF) and through the Basic Oxygen Steelmaking (BOS) process.

The key component in the BOS is the Basic Oxygen Converter, however before this process can begin a blast furnace is required to create a charge of molten iron. The raw materials for producing molten iron are iron ore, coking coal and fluxes (materials that help the chemical process) - mainly limestone. Blended coal is first heated in coke ovens to produce coke (carbonisation), after which it is allowed to cool. Iron ore lumps and pellets, coke, sinter and possibly extra flux are carried to the top into the blast furnace. Hot air (900°C) is blasted into the bottom of the furnace, from which oxygen combusts with the coke forming CO, which flows up through the blast furnace, removing oxygen from the iron ores on their way down, thereby leaving iron. The heat in the furnace melts the iron, and the resulting liquid iron flows out at the bottom of the furnace, towards the BOS vessel in which scrap steel has been charged first. Then very pure oxygen is blown at high pressure, which combines with the carbon, separating them from the metal, leaving steel.

Unlike BOS, the EAF is charged with "cold" material (recycled steel goods at end-of-life, or direct reduced iron (DRI) and iron carbide, as well as pig iron). The cold material is fed into the furnace, after which electrodes are lowered into it. An electric current is passed through the electrodes to form an arc. The heat generated by this arc melts the scrap. As with the basic oxygen process, oxygen is blown in to the furnace to purify the steel.

# The UK's three integrated blast furnace sites have multiple sources of $CO_2$ emissions




Sites are also heterogeneous: For example, the Teesside site has one large blast furnace, whereas the Scunthorpe site has four smaller blast furnaces (of which two are currently in operation). This has implications for scale and technology that require site process modelling.

Source: Carpenter, A. (2012)

# Iron and Steel in the UK– $CO_2$ sources, sites and barriers

## UK sites

Site $CO_2$ emissions between 2008 and 2012 for the three integrated steelworks was between 6.2 to 7.3 $MtCO_2$/yr per site, making these sites the largest single industrial $CO_2$ emitters.

## $CO_2$ sources and emissions

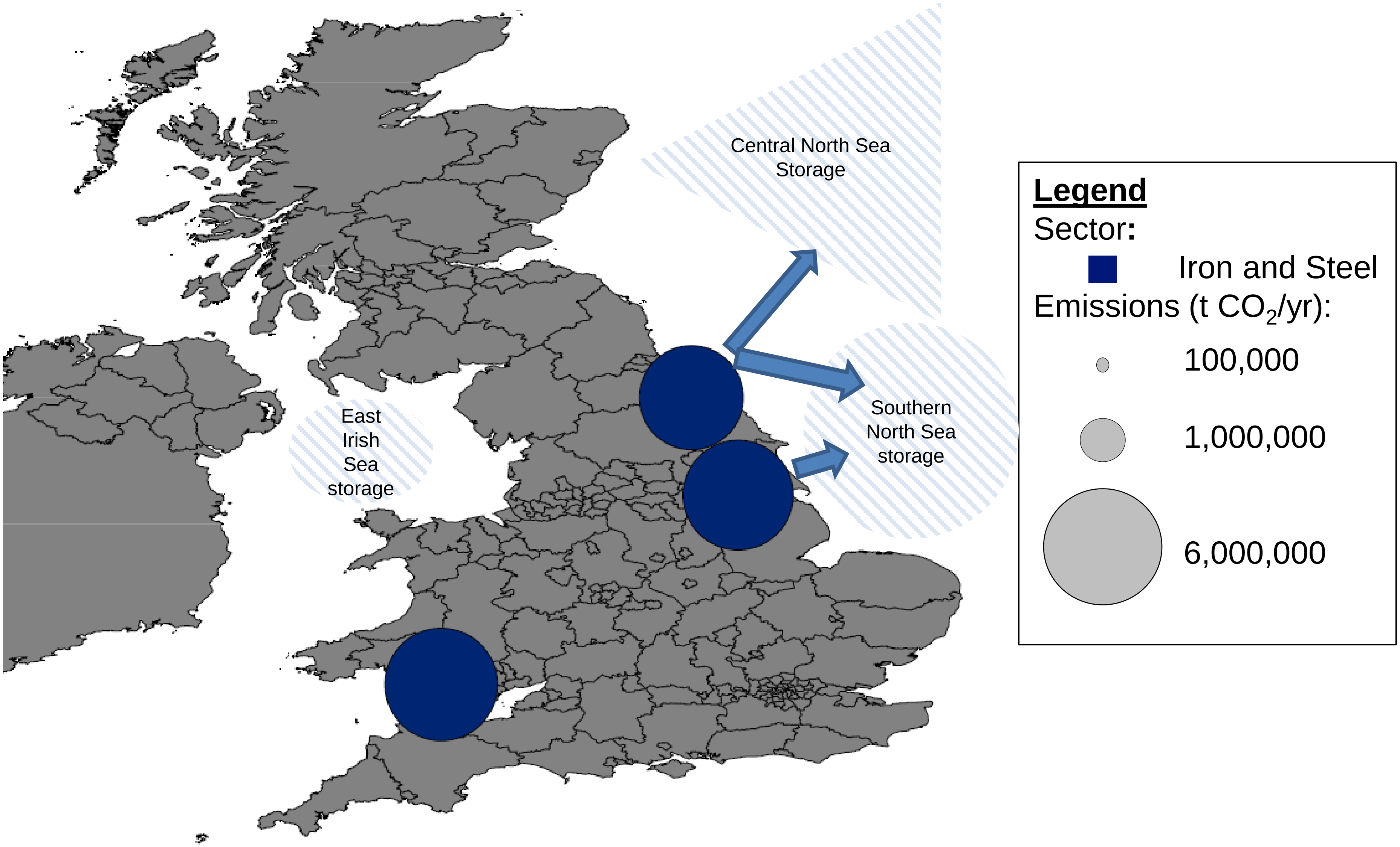
Main sources of $CO_2$ emissions in steel production are the blast furnace, iron ore reduction and sintering plant, and coke production.

| CO2 emission point | Description | % of total emissions | % concentration of $CO_2$ stream | Notes |
|---|---|---|---|---|
| Blast Furnace (BF) | Primary process to produce iron by heating coke, pulverised coal, sinter and bulk ore to ~1,500°C | ~69% | 16% - 26% [Remus, R., & Roudier, S. (2010). ] | Flue gas from BF used for power generation |
| Sinter | For iron production | 16% | 5-10% | |
| Coke plant | For iron production | 16% | 25% | |

# Iron and steel - technology barriers for deploying a capture project in the period 2020 to 2025

| Technology | Barriers | |
|---|---|---|
| **General** | • Total site emissions of iron and steel plants, 4-8 Mt/yr, are very large compared to the other sectors. Capture of 100% of site $CO_2$ emissions not realistic for projects taking FID in 2020 for operation in 2025, but there are multiple options for partial capture. | • Multiple processes and vents in an integrated site – challenge to identify "optimal" CCS solutions.<br>• Need to synchronise integration with a major overhaul, which have intervals longer than 7 years, so few and limited windows of opportunity, without incurring significant additional downtime costs. |
| **1st generation chemical solvents (MEA/MDEA)** | • Risk of lock-in to a technology option with high system cost.<br>• Little work on direct application to BFG, therefore compatibility unclear, (although should be easier to apply to the CHP system).<br>• Need for steam to drive capture plant. Limited familiarity of UK sites with amine-based capture, except in NOx and SOx control. | • Contaminants in flue gas results in degradation (NOx) and salt formation (SOx), resulting in a high amine requirement. |
| **Physical solvents** | • Assuming configuration involves application of physical solvent immediately after Blast Furnace, there is a lack of experience in compressing and cooling Blast Furnace Gas (which is hot and is a mixture of several reactive chemicals), for use with a physical solvent. CO passes through, implying upper limit to the fraction of $CO_2$ that can be captured. | • Will be a need to reconfigure optimal site energy and flow balances to account for difference in product stream pressure, temperature and composition reaching the CHP system.<br>• Risk of "all-or-nothing" process for each Blast Furnace, with lock-in of BFG composition and hence BF process. |
| **2nd generation chemical solvents (e.g. advanced amines, amino acids and blends)** | • Limited effort targeted at developing solvents which are compatible and efficient with $CO_2$ streams from iron and steel sites. | • Unfamiliarity with using ammonia poses an operational/safety barrier to implementing an ammonia option.<br>• Similar challenges as for first generation amines, although expected to be less challenging. |
| **Solid looping e.g. Calcium looping** | • TRL6 technology requiring significant piloting and demonstration before it can be applied at scale of 1 Mt/yr or higher that is relevant for steel plants.<br>• Not clear if by-product salts can be sold. | • Available performance models need refining. Extent of calcium looping integration with core process is unclear – ideally would use $CO_2$ rich flue gas at high temperature directly rather than cooled streams. |
| **Top gas recycling** | • Fundamental analysis and redesign of Blast Furnace required to cope with changes in mass and energy flows associated with TGR-BF. Would need to time with major overhaul. | • Risk of project-on-project risk.<br>• Current investigation still mainly focused on carbon separation, not yet at end to end projects. |
| **Oxyfuel technologies** | • Potential for application to stoves but this represents less than a quarter of site emissions so there is a limited maximum capture potential per site. | • Standard oxyfuel challenges associated with compatibility of existing materials and heat flow patterns with flame temperatures from oxyfuel. |

# If $CO_2$ transport networks are developed in Teesside and Yorkshire, these could serve SSI and Tata Scunthorpe sites respectively.



# Iron and Steel $CO_2$ Source Database Assumptions

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2$/yr) | COMAH |
|---|---|---|---|---|---|---|---|
| Iron and Steel | Tata Steel UK Limited | Scunthorpe Integrated Iron & Steel Works | Steel | 53.57 | - 0.61 | 7,305,903 | Top |
| Iron and Steel | Tata Steel UK Limited | Port Talbot Steelworks | Steel | 51.56 | - 3.77 | 6,880,337 | Top |
| Iron and Steel | SSI | Teesside Integrated Iron & Steel Works | Steel | 54.61 | - 1.11 | 6,222,710 | Top |

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# Source sectors

- Cement
- Chemicals
- Iron and Steel
- Oil Refining
- Other sectors of potential relevance

# Refineries – Sector description

## Sector context

The members of UKPIA (UK Petroleum Industry Association) run the seven major operating refineries in the UK, which are responsible for more than 90% of the $CO_2$ emissions from the sector. Since the refinery closures in 1997, 1999, 2009, and most recently 2012, UK refining throughput has fallen from its late 90s' peak of 97 million tonnes of crude oil to around 69 million tonnes in 2012 – an 8% drop compared to 2011. Over 80% of product output is petrol, diesel, jet fuel, gas oil and fuel oils.

## Description of main processes

Refinery operations can be broken down into five main processes: (1) distillation (separates crude oil into different refinery streams) – (2+3) conversion and reforming (quality improvement and yield adjustments to meet market demand) – (4) desulphurisation (reduces sulphur in the streams) – (5) blending of the refinery streams (to produce final products).

The starting point for all refinery operations is the crude distillation unit (CDU). Crude oil is boiled in a fractioning column, which breaks the crude down into more useful components. The crude oil enters the column near the bottom and is heated to around 380°C. The lighter fractions are vaporised and rise up the column. As they rise, they are cooled by a downward flow of liquid and condense at different points. This enables fractions with different boiling points to be drawn off at different levels in the column.

High purity $CO_2$ is co-produced as a by-product of hydrogen production at the Lindsay oil refinery.

# Refineries in the UK– $CO_2$ sources, sites and barriers

## $CO_2$ sources and emissions

Refineries emit around 30% of the UK's industrial $CO_2$ emissions and are included in the EU ETS. The total emissions for the last 20 years account to 15 – 20 $MtCO_2e$/y. There are four major $CO_2$ emission sources in a refinery: furnaces and boilers, utilities, fluid catalytic cracker (FCC) and hydrogen production.

| $CO_2$ emission point | Description | % of total refinery emissions | | % concentration of CO2 stream [OECD, IEA, & UNIDO, 2011] |
|---|---|---|---|---|
| | | UKPIA | OECD, et all (2011). | |
| FCC (post clean up) | Process used to convert crude oil to more valuable products. | 13-32% | 20-50% | 10-20% |
| Utilities | $CO_2$ from production of electricity and steam on site | 12-27% | 20-50% | 3-6% |
| Furnaces | Heat is required for reactions in the refining process such as cracking, reforming and steam generation. | 12-27% | 30-60% | 8-10% |
| Hydrogen | Refineries require hydrogen for numerous processes. Not all refineries produce hydrogen on-site. | 11% | 5-20% | 20-99% |

# Refineries


*Image kindly provided by Ineos showing the very large number of potential vents at the Grangemouth refinery-chemical-CHP complex.*

# Refining - carbon capture technology barriers prioritised in stakeholder interviews.

| Technology | Barriers |
| --- | --- |
| **1st generation chemical solvents** | • Contaminants in flue gas results in degradation (NOx) and salt formation (SOx), resulting in a high amine requirement |
| **Chilled ammonia** | • Usually no ammonia on site, would introduce new risks |

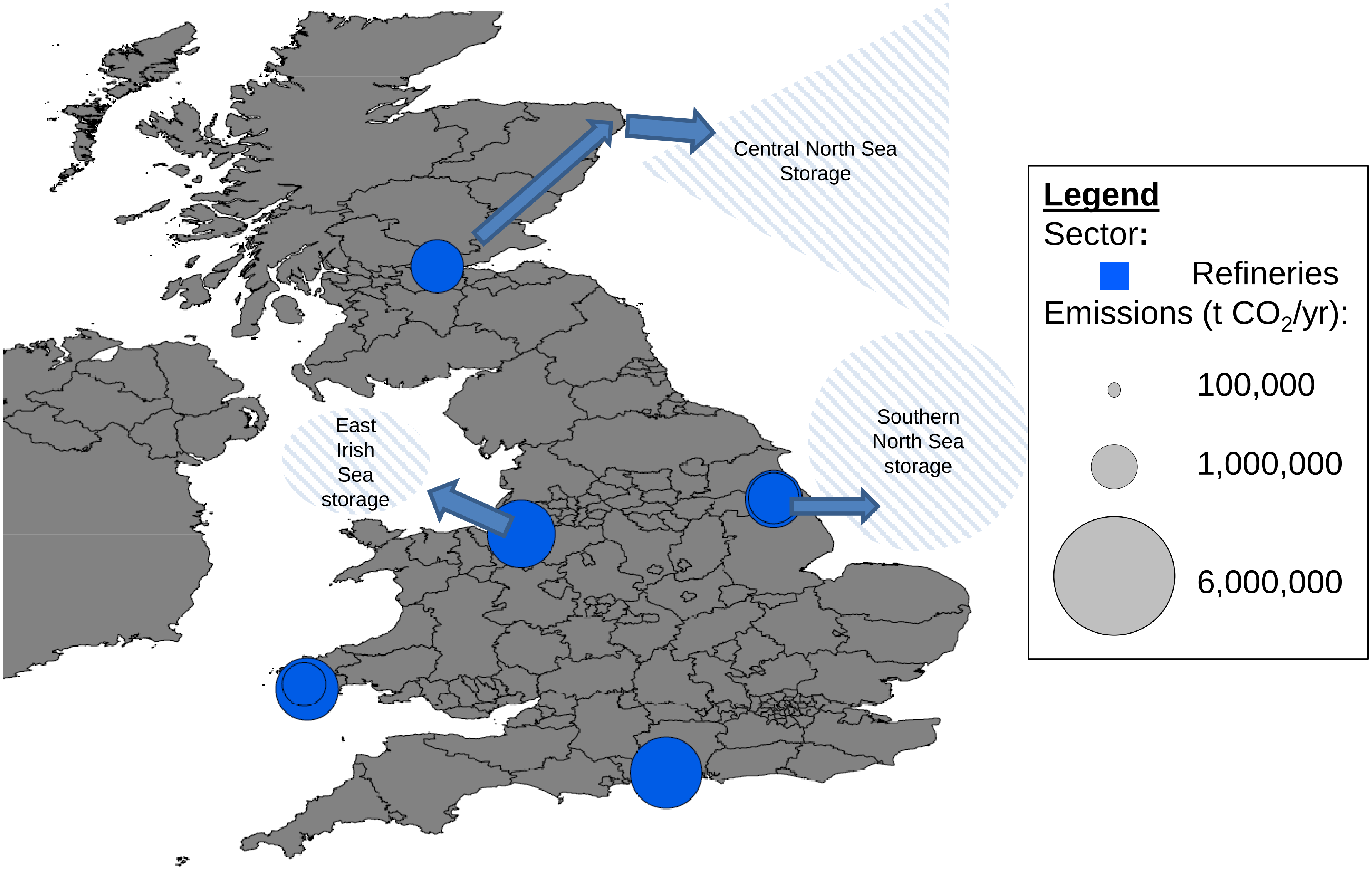
- This list of barriers is not comprehensive, but reflects barriers prioritised during stakeholder interviews.

# Refining - sector specific barriers

| Issue | Barriers |
|---|---|
| **Major overhaul turn around time** | • Turn around time between major overhauls is five years, this can limit speed of development |
| **Furnace shut down operation issues** | • Operating issues when shutting down a furnace in a $CO_2$ multi unit carbon capture facility.<br>• Significant downtime costs, potential impacts of £M/day |
| **Multiple source exhaust** | • Many distributed emission points across a site. Either large duct network at significant investment and blower operational cost as well as integration challenges or high investment in multiple capture plants. |
| **Cost uncertainty** | • Experience suggests cost uncertainties of at least +200%/-33% are common in this sector |

Feedback from A. Roberts, UKPIA, Personal Communication

# Transport and storage opportunities most likely for refineries in Scotland, NW England and Yorkshire



# Refining Sector $CO_2$ Source Database Assumptions (1/2)

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2$/yr) | COMAH |
|---|---|---|---|---|---|---|---|
| Refining | Esso Petroleum Company Limited | Fawley refinery, Southampton1 | Refinery | 50.83 | - 1.35 | 1,297,000 | Top |
| Refining | Esso Petroleum Company Limited | Fawley refinery, Southampton2 | Petrochemical cracker (olefins) | 50.83 | - 1.35 | 800,000 | Top |
| Refining | Esso Petroleum Company Limited | Fawley refinery, Southampton3 | Industrial Gas CHP | 50.83 | - 1.35 | 520,000 | Top |
| Refining | Ineos Manufacturing Scotland Ltd | Grangemouth Refinery cracker | Petrochemical cracker (olefins) | 56.01 | - 3.70 | 350,000 | Top |
| Refining | Ineos Manufacturing Scotland Ltd | Grangemouth CHP | Industrial Gas CHP | 56.01 | - 3.70 | 723,000 | Top |
| Refining | Ineos Manufacturing Scotland Ltd | Grangemouth Refinery excl. cracker and CHP | Refinery | 56.01 | - 3.70 | 1,622,000 | Top |
| Refining | Essar Oil UK Ltd | Stanlow Refinery FCC | Petrochemical cracker (olefins) | 53.27 | - 2.84 | 600,000 | Top |
| Refining | Essar Oil UK Ltd | Stanlow Refinery Power generation | Refinery | 53.27 | - 2.84 | 450,000 | Top |
| Refining | Essar Oil UK Ltd | Stanlow Refinery excl FCC and power plant | Refinery | 53.27 | - 2.84 | 1,630,000 | Top |
| Refining | Phillips 66 (formerly Conoco Phillips) Limited | Humber Refinery FCC regenerator stack1 | Petrochemical cracker (olefins) | 53.63 | - 0.25 | 450,000 | Top |
| Refining | Phillips 66 (formerly Conoco Phillips) Limited | Humber Refinery FCC regenerator stack2 | Refinery | 53.63 | - 0.25 | 1,499,000 | Top |

# Refining Sector $CO_2$ Source Database Assumptions (2/2)

| Sector | Owner name as listed in the ETS | Source name | Source archetype | Latitude | Longitude | $CO_2$ emissions ($tCO_2$/yr) | COMAH |
|---|---|---|---|---|---|---|---|
| Refining | Valero Energy Ltd | Pembroke Refinery FCC regenerator stack | Petrochemical cracker (olefins) | 51.69 | - 5.03 | 800,000 | Top |
| Refining | Valero Energy Ltd | Pembroke Refinery excl. FCC | Refinery | 51.69 | - 5.03 | 1,475,000 | Top |
| Refining | Murco Petroleum Limited | Murco Petroleum Milford Haven Refinery1 | Refinery | 51.74 | - 5.06 | 780,000 | Top |
| Refining | Murco Petroleum Limited | Murco Petroleum Milford Haven Refinery2 | Petrochemical cracker (olefins) | 51.74 | - 5.06 | 300,000 | Top |
| Refining | Murco Petroleum Limited | Murco Petroleum Milford Haven Refinery3 | Industrial Gas CHP | 51.74 | - 5.06 | 300,000 | Top |
| Refining | Lindsey Oil Refinery | Total Lindsey Oil Refinery Hydrogen plant | Hydrogen via SMR | 53.64 | - 0.25 | 200,000 | Top |
| Refining | Lindsey Oil Refinery | Total Lindsey Oil Refinery CHP Plant | Industrial Gas CHP | 53.64 | - 0.25 | 300,000 | Top |
| Refining | Total UK Limited | Total Lindsey Oil Refinery FCC regenerator stack | Refinery | 53.64 | - 0.25 | 450,000 | Top |
| Refining | Total UK Limited | Total Lindsey Oil Refinery Other | Refinery | 53.64 | - 0.25 | 669,000 | Top |

# References – Oil Refining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# Source sectors

- Cement
- Chemicals
- Iron and Steel
- Oil Refining
- Other sectors of potential relevance

# Other sources of $CO_2$ of potential relevance to industrial CCS

Although the scope of this study is limited to existing large UK cement, chemicals, iron and steel and oil refining sites, policymakers should be aware that other sources of $CO_2$ that may be relevant as part of wider industrial CCS or CCU development over the long-term include:

- Many industrial sites have one or more existing boilers, furnaces and/or CHP units, which may use one or more of coal, gas, oil, biomass, or waste fuel.
- New build industrial sites, especially if built “capture ready”.
- Other heat-intensive industrial sectors, such as glass, ceramics, pulp and paper, food and drink.
- Hydrocarbon processing (e.g. $CO_2$ is coproduced with oil or gas and separated at offshore installations and at St. Fergus gas terminal), often generating $CO_2$ rich vents. (It is not clear if there will be $CO_2$ co-produced with any potential future UK shale gas production.)
- Biofuel production, for example bio-ethanol production through fermentation.
- Power stations fitted with pre-combustion capture could supply “low carbon” hydrogen to industrial sources.

# Source archetype assumptions

| Source archetype | Vent complexity | CO2 stream pressure (MPa) | Input CO2 (% volume) | | | Impurity level (ppm) | | %CO2 site capturable |
|---|---|---|---|---|---|---|---|---|
| | | | Low | Central | High | NOx | SOx | |
| Steel | Many | 0.11 | 16% | 30% | 44% | 100 | 100 | 60% |
| Cement | Single | 0.11 | 14% | 24% | 33% | 100 | 100 | 99% |
| Other Refinery | Many | 0.11 | 8% | 10% | 12% | 600 | 1200 | 90% |
| Hydrogen | Single | 0.11 | 20% | 95% | 99% | 10 | 10 | 99% |
| Gas boiler | Single | 0.11 | 5% | 7% | 10% | 10 | 10 | 99% |
| Industrial Coal CHP | Single | 0.11 | 8% | 12% | 15% | 100 | 100 | 99% |
| Industrial Gas CHP | Single | 0.11 | 2% | 3% | 6% | 10 | 10 | 99% |
| Ammonia | Single | 0.1-5.1 | 90% | 95% | 100 % | 100 | 100 | 99% |
| Crackers | Many | 0.11 | 8% | 10% | 12% | 100 | 100 | 99% |
| Other chemicals | Many | 0.11 | 2% | 11% | 40% | 100 | 100 | 99% |

# Outline

- Overall Project Methodology
- $CO_2$ capture technologies
- $CO_2$ sources
- Techno-economic analysis of industrial $CO_2$ capture
- Process simulation case studies
- $CO_2$ utilisation review

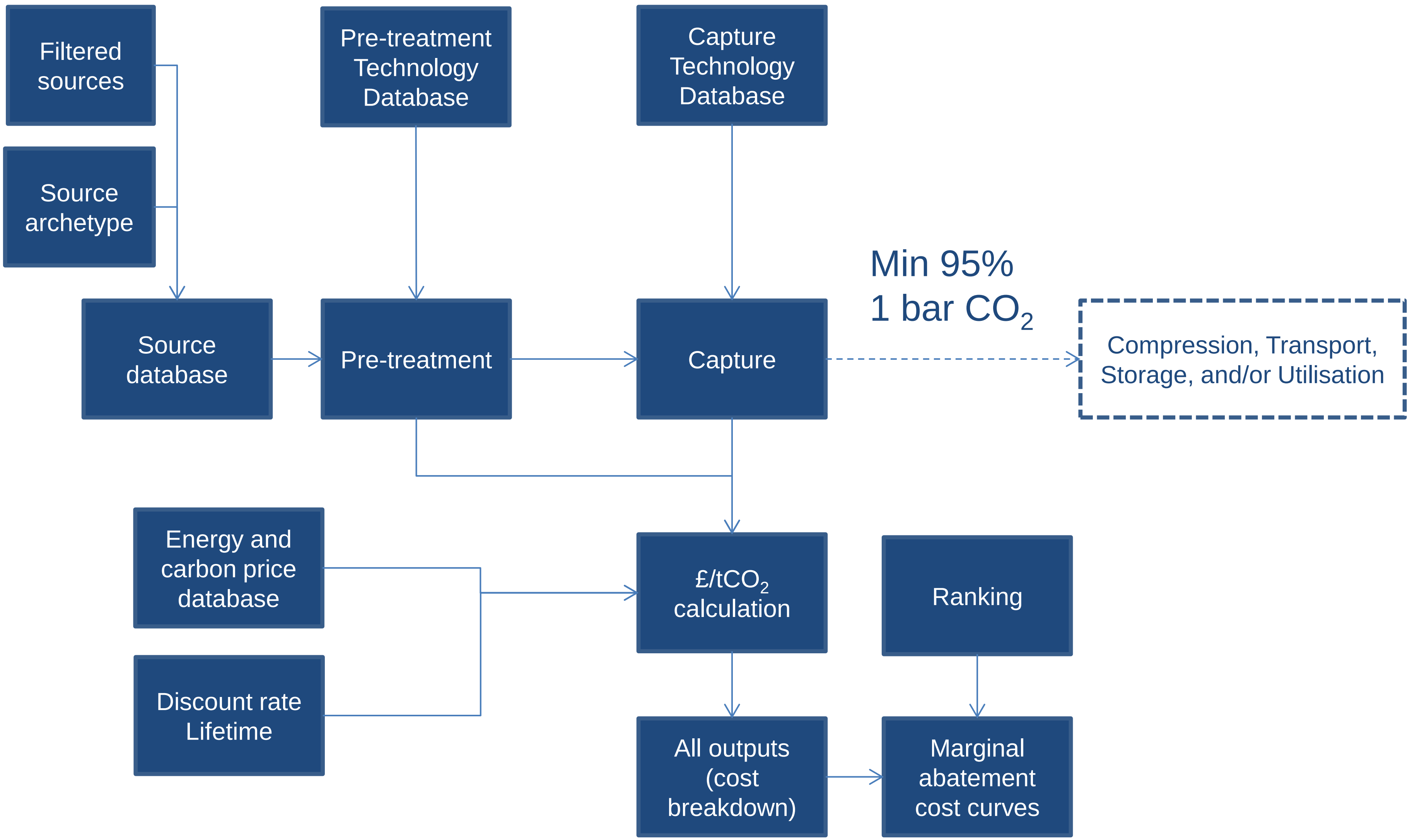
# Techno-economic analysis of industrial $CO_2$ capture

- Model Architecture
- Key assumptions
- Pragmatic scenario
- Sensitivity analysis

# Techno-economic modelling is used to screen the potential of $CO_2$ capture technologies for each industrial sector.

- The model “CINDY V1.0” (short for Capture in INDustrY) was developed by Element Energy to facilitate rapid Excel-based screening of source-technology combinations across a range of user-defined scenarios.
- Key technology selection and performance indicators are availability, costs and overall $CO_2$ saving potential. The range of costs (in £/t$CO_2$ captured or abated) can be estimated for each capture technology from capex, opex, heat and power consumption.
- The published literature is heterogeneous in the level of detail and consistency of input, output and process assumptions. There are many drivers of costs however; therefore it is a major challenge to compare the published costs for industrial $CO_2$ capture.
- The model allows a more systematic comparison of sectors and technologies, although significant uncertainties and assumptions are still required.

# Architecture for techno-economic model



# Summary description of the Excel-based techno-economic model

- Filtered source database, scaling sub-sector archetype properties with actual source $CO_2$ emissions and local cost of heat.
- Costs and energy requirements disaggregated using initial pre-treatment, capture technology,  and post-treatment.
- Initial pre-treatment and post-treatment steps reflect high level models, whereas the capture technology "archetype" combines reference data on suitability (with a given $CO_2$ stream), cost and performance with correlation functions to allow costs and performance to be adjusted based on key drivers ($tCO_2$/yr, input purity, and %$CO_2$ captured).
- Once input conditions are selected, the model cycles through calculations of capture costs and $CO_2$ savings for all relevant source-technology combinations (initially for 2020 and subsequently for 2025). From these initial outputs, the technology-source combinations can then be ranked, based on annual $CO_2$ captured or abated, annualised £/$tCO_2$ captured or abated.
- An overall technology assessment matrix can then be determined identifying for each of the four sectors, sub-total $CO_2$ captured or abated, average £/$tCO_2$ captured or abated.
- Users can carry out sensitivity analysis to discount rate, energy price, technology and source assumptions.

# Deriving the cost of capture

Capital costs = Pre-treatment cost + Capture cost

Compression, transport, storage are excluded from baseline analysis.

£/t$CO_2$ abated costs will be presented as levelised costs, i.e. using a discounted cashflow approach:

$$Levelised\ cost\ of\ capture\ \left(\frac{£}{tCO_2}\right) = \frac{PV(capex) + PV\ (fixed\ opex) + PV\ (heat) + PV\ (elec)}{PV\ (lifetime\ abated\ tCO_2)}$$

To obtain the present values (PV), the capex, opex and abated t$CO_2$ will be discounted at the user-defined discount rate recognising construction period (default equally spread over 3 yrs) and user-defined project operational lifetime (default is 15 yrs).

$CO_2$ and £/t$CO_2$ will be provided as both captured and abated.

Within the discounted cashflow costs and $CO_2$ are discounted to the first year of operation (e.g. 2025).

# Techno-economic analysis

- Model Architecture
- Key assumptions
- Pragmatic scenario
- Sensitivity analysis

# Key baseline assumptions

- Baseline conditions were agreed with the project steering board at an interim meeting as follows:
  - Real discount rate 10% (costs discounted to date that project begins operation e.g. 2025)
    - A discount rate of 10% is used across the CCS literature, and is used to facilitate comparison. It is likely that any industry investors would demand higher hurdle rates. Real discount rate implies no additional adjustments required for inflation.
  - Project lifetime 15 yrs
  - DECC Central Prices for electricity, gas, carbon
  - Construction period 3 yrs
  - Costs standardised to £(2013) where possible.
  - Source “central” % $CO_2$ concentration
  - Source “central” $MtCO_2$/yr emissions
  - No waste heat recovery and re-use assumed for initial projects (projects assumed to include cost of a boiler where significant heat demand is required).
  - No consideration of tax
  - Costs exclude post-capture compression, transport or storage
  - Assume for techno-economic modelling only one capture technology, i.e. combinations of capture techs are excluded.
    - General principles on the design of capture using multiple technologies are not well described in the industrial CCS literature.
    - For examples of approaches combining solvent absorption, pressure swing adsorption and/or cryogenic $CO_2$ capture in the iron and steel sector see IEA GHG (2013) Iron and Steel CCS Study (Techno-economics integrated steel mill) Report 2013/04.

# DECC Energy and Carbon Price Assumptions Used

| | Electricity p/kWh | | | Gas p/kWh | | | Traded $CO_2$ £/t | | | Non traded $CO_2$ £/t | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | Low | Central | High | Low | Central | High | Low | Central | High | Low | Central | High |
| 2020 | 10.23 | 12.02 | 13.77 | 1.91 | 3.13 | 4.26 | - | 4.87 | 25.98 | 32.85 | 65.71 | 98.56 |
| 2021 | 10.68 | 12.34 | 13.96 | 1.92 | 3.13 | 4.35 | 3.81 | 12.01 | 34.82 | 33.4 | 66.8 | 100.2 |
| 2022 | 11.06 | 12.77 | 14.75 | 1.92 | 3.14 | 4.36 | 7.62 | 19.14 | 43.65 | 33.95 | 67.9 | 101.84 |
| 2023 | 11.38 | 13.02 | 14.52 | 1.93 | 3.15 | 4.37 | 11.43 | 26.28 | 52.49 | 34.5 | 68.99 | 103.49 |
| 2024 | 12.16 | 13.56 | 15.24 | 1.94 | 3.16 | 4.38 | 15.25 | 33.41 | 61.33 | 35.04 | 70.09 | 105.13 |
| 2025 | 12.14 | 13.63 | 15.03 | 1.95 | 3.17 | 4.39 | 19.06 | 40.55 | 70.16 | 35.59 | 71.18 | 106.77 |
| 2026 | 12.69 | 13.79 | 15.19 | 1.96 | 3.18 | 4.39 | 22.87 | 47.69 | 79 | 36.14 | 72.28 | 108.41 |
| 2027 | 12.40 | 13.72 | 15.06 | 1.96 | 3.18 | 4.4 | 26.68 | 54.82 | 87.84 | 36.69 | 73.37 | 110.06 |
| 2028 | 12.34 | 13.6 | 15.37 | 1.97 | 3.19 | 4.41 | 30.49 | 61.96 | 96.67 | 37.23 | 74.47 | 111.7 |
| 2029 | 12.27 | 13.52 | 14.77 | 1.98 | 3.2 | 4.42 | 34.3 | 69.1 | 105.51 | 37.78 | 75.56 | 113.34 |
| 2030 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 38.12 | 76.23 | 114.35 | 38.12 | 76.23 | 114.35 |
| 2031 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 41.89 | 83.77 | 125.66 | 41.89 | 83.77 | 125.66 |
| 2032 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 45.45 | 90.89 | 136.34 | 45.45 | 90.89 | 136.34 |
| 2033 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 49.01 | 98.01 | 147.02 | 49.01 | 98.01 | 147.02 |
| 2034 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 52.56 | 105.13 | 157.69 | 52.56 | 105.13 | 157.69 |
| 2035 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 56.12 | 112.25 | 168.37 | 56.12 | 112.25 | 168.37 |
| 2036 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 59.68 | 119.36 | 179.05 | 59.68 | 119.36 | 179.05 |
| 2037 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 63.24 | 126.48 | 189.72 | 63.24 | 126.48 | 189.72 |
| 2038 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 66.8 | 133.6 | 200.4 | 66.8 | 133.6 | 200.4 |
| 2039 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 70.36 | 140.72 | 211.08 | 70.36 | 140.72 | 211.08 |
| 2040 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 73.92 | 147.84 | 221.75 | 73.92 | 147.84 | 221.75 |
| 2041 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 77.48 | 154.95 | 232.43 | 77.48 | 154.95 | 232.43 |
| 2042 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 81.04 | 162.07 | 243.11 | 81.04 | 162.07 | 243.11 |
| 2043 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 84.6 | 169.19 | 253.79 | 84.6 | 169.19 | 253.79 |
| 2044 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 88.15 | 176.31 | 264.46 | 88.15 | 176.31 | 264.46 |
| 2045 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 91.71 | 183.43 | 275.14 | 91.71 | 183.43 | 275.14 |
| 2046 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 95.27 | 190.54 | 285.82 | 95.27 | 190.54 | 285.82 |
| 2047 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 98.83 | 197.66 | 296.49 | 98.83 | 197.66 | 296.49 |
| 2048 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 102.39 | 204.78 | 307.17 | 102.39 | 204.78 | 307.17 |
| 2049 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 105.95 | 211.9 | 317.85 | 105.95 | 211.9 | 317.85 |
| 2050 | 12.36 | 13.47 | 15.11 | 1.99 | 3.21 | 4.43 | 109.51 | 219.02 | 328.53 | 109.51 | 219.02 | 328.53 |

Table shows undiscounted costs.

# Pre-treatment assumptions (1): Pipeline gathering network

- A simple model for pipeline gathering was assumed.
- Assumes streams are compatible. (Streams not being comparible might result in the need for multiple independent capture trains, or more likely, less $CO_2$ captured).
- Capital cost =f(throughput, no. of vents),
- Throughput and no. of vents are estimated relative to a reference project, assuming
- The number of vents is "single (1)", "few (2-4)" or "many (>5)" vents),
- The reference project is the Simmonds *et al*. (2003) study of Grangemouth oil refinery.
- Costs have been updated to £2013 to account for inflation.

**Vent collection**

| Type | Capex (£m) | Reference $CO_2$ (Mt captured/y) | Reference $CO_2$ purity (% volume) |
|---|---|---|---|
| Single | £- | 2 | 10.0% |
| Few | £ 15 | 2 | 10.0% |
| Many | £ 60 | 2 | 10.0% |

## Pre-treatment costs (2): Impurity removal

- Some capture technologies are vulnerable to impurities in the flue gas stream.
- For these scenarios, a simple cost model is assumed to reduce the levels of $NO_x$ and $SO_x$ based on conventional technologies used in the power and industrial sectors (e.g. Selective Catalytic Removal, Flue Gas Desulfurisation).
- No assumption is made on the ability to sell the products of $NO_X$/$SO_X$ removal.
- Some capture technologies (e.g. physical solvents) are only relevant when the input $CO_2$ is at high partial pressure. Given the higher cost of electricity than heat, it is unusual to increase the pressure of the flue gas and then use one of these technologies, however it is feasible.
- Therefore the pre-treatment for the combination of low pressure gases with capture technologies requiring high pressure $CO_2$ may therefore include the costs of compression.  This would be based on the sizing and costs of conventional compressors.

# Pre-treatment assumptions

Four high level and simplified “pre-treatment” techno-economic models are included, depending on source-technology compatibility requirements.

- $SO_X$ removal (costs in line with FGD)
- $NO_X$ removal (costs in line with SCR)
- Costs are scaled in line with flue gas throughput relative to a reference value.
- Scaling algorithm is based on standard engineering rule of thumb (based on surface area: volume relationship)

$$cost_A/cost_B=(scale_A/scale_B)^{2/3}$$

**Flue impurity processing**

| | | | | Impurity tolerance removal index | | | |
|---|---|---|---|---|---|---|---|
| Technology | Name | Reference capex (£m 2013) | Reference flue (m3/y) | 50% | 95% | 99% | Fixed opex (% of capex in 2013) |
| FGD | Flue gas desulphurisation | £ 99 | 20,000,000,000 | 50% | 100% | 200% | 5% |
| SCR | Selective catalytic reduction | £ 99 | 20,000,000,000 | 50% | 100% | 200% | 5% |

# References for flue gas clean-up


- DTI (2000) Flue Gas Desulphurisation (FGD) Technologies *DTI Cleaner Coal Programme Technology Status Report 012*
- Markusson, N. (2012) Scaling up and deployment of FGD in the US (1960s to 2009) UKERC /RS/CCS/2012/006 (Final case study report as part of Work Package 2 of the UKERC project: CCS – Releasing the Potential?)
- Cichanowicz, J.E. (2010) Current capital cost and cost-effectiveness of power plant emissions control technologies, prepared for the Utility Air Regulatory Group
- Environment Agency TWG12 (2011) Best Available Technology for SO2 for existing baseload UK Coal Units > 300 MW downloaded from http://www.environment-agency.gov.uk/static/documents/Business/UKTWG12_Final_SO2_baseload_coal.pdf
- Environment Agency TWG12 (2011) Best Available Technology for NOx for existing baseload UK Coal Units > 300 MW downloaded from http://www.environment-agency.gov.uk/static/documents/Business/UKTWG13_Final_NOx_baseload_coal.pdf
- US Environmental Protection Agency Air Pollution Control Technology Factsheet EPA-452/F-03-034

# Pre-treatment (3): Flue gas compression

- Gas compression is a mature technology which is not the focus of the present study, therefore a simplified high level model is used for compression.
- Assume NOx and SOx removal (which results in gas with a starting pressure of 1 bar) and pipeline gathering occurs prior to flue gas compression.
- Assume flue gas compression from 1 bar to the minimum required for capture technology.
- Assume that flue gas is predominantly $N_2$ and $CO_2$, and can be compressed adiabatically.

$$[(P_{out}/P_{in})^\wedge\gamma-1]/\gamma = 5.15 \text{ for } CO_2$$

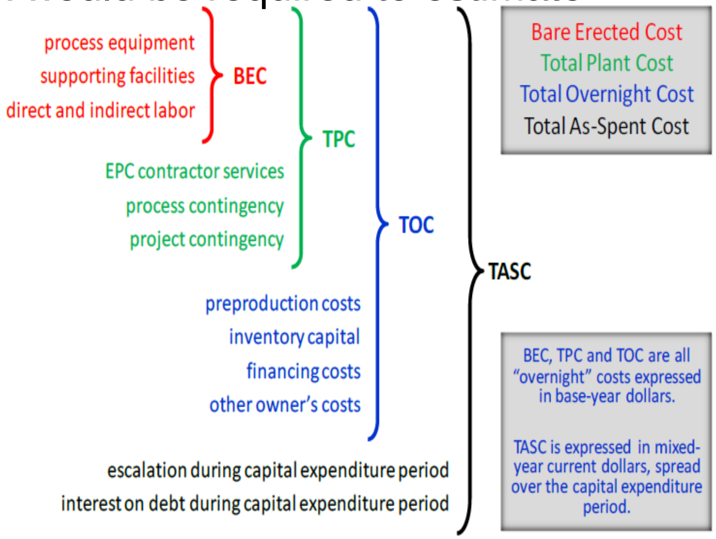
$$[(P_{out}/P_{in})^\wedge\gamma-1]/\gamma = 5.79 \text{ for } N_2$$

- Assume an efficiency of 75%
- Assume compressor capital cost of ca. £1.64m/MW
  - Updated from estimate of £1,238/kW in 2005
  - Assume fixed annual operating costs of 5% of capex.
  - Electricity cost at relevant industrial electricity price.

McCollum and Ogden (2006) Techno-economic Models for CO2 Compression, Transport, and Storage, Institute for Transportation Studies, UCDavis (UCD-ITS-RR-06-14)

# Basis of plant cost

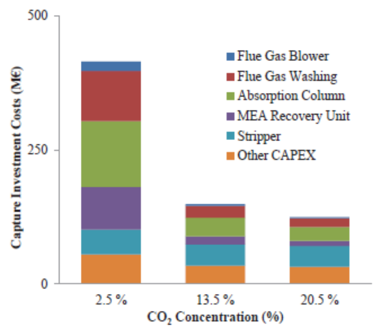
- As this project has a focus on technology and site comparison, this project compares technologies at the level of Total Plant Cost (see below).
- Costs for techno-economic model were developed from the literature review (updated where necessary following discussions with experts or for the purposes of standardising conditions).
- Given the paucity of realised industrial CCS projects, and the diversity of transparency, scope, date, and assumptions in the literature (previously described in Element Energy *et al*. 2013 for BIS), standardisation of costs is very challenging.
- The model does allow contingencies, owner's costs, and financing to be analysed through sensitivity analysis. Note that contingencies and financing costs for first-of-a-kind project may be significant, although further work would be required to estimate how these may vary between technologies or sites.
- An independent approach is used for the process simulation work described later in this appendix.

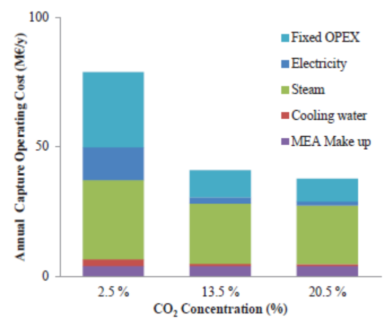



Element Energy (2013) CCS in UK industry in 2030 - A High level cost review, for BIS
NETL 2010 Costs and Performance of Fossil Fuel Power Plants;
Kuramochi et al. Techno-economics of industrial CCS

# Capital cost adjustment from a reference data source

- Cost reduction due to economies of scale assumed to follow an empirical “2:3 power law” (resulting conceptually from the ratio of changing surface areas, which drive cost,  and volume, which drives capacity).
- The cost adjustment factor for scale is given by $cost_B=cost_{ref}*(Scale_B/Scale_{ref})^{0.66}$
- Analysis by Husebye *et al.* (2012) also identifies the potential for significant reduction in capital (below left) and operating (below right) costs for MEA capture as a function of flue gas $CO_2$ concentration.

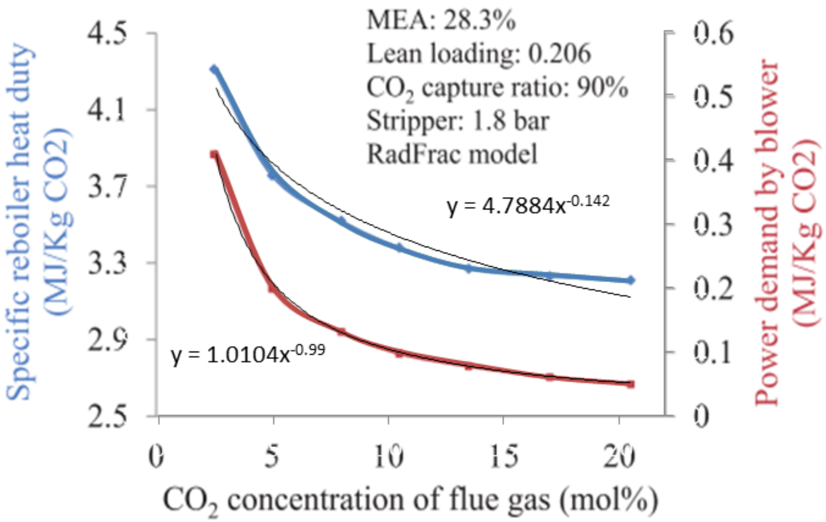


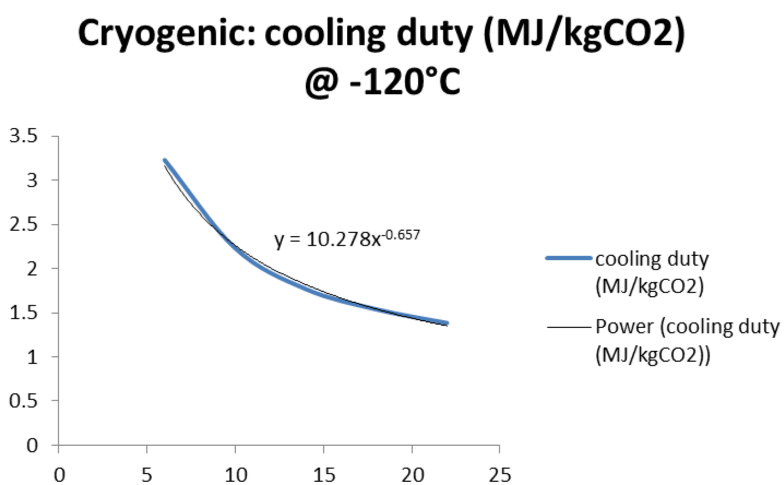




- Much of the capex reduction on increasing $CO_2$ concentration can be attributed to reduced diameters (at constant flue gas throughput). We assume that the cost dependency identified for MEA holds for other 1st generation amines and all other capture technologies identified, as it relates to fundamental equilibrium processes, although this would ideally be validated through demonstration projects.
- Capital costs were adjusted for $CO_2$ concentration through
  - Capital cost(scale, [%$CO_2$]) = reference cost x cost adjustment factor (scale) x cost adjustment factor for [%$CO_2$], where the cost adjustment factor for [%$CO_2$] is given by $312.5*(2.5/[\%CO_2])^{0.53}$

Husebye, J. *et al*. (2012) Techno-economic evaluation of amine-based CO2 capture: impact of CO2 concentration and steam supply *Energy Procedia* **23** 381-390

# Assumptions on adjustment of energy demand for capture as a function of $CO_2$ concentration




There are only a handful of data points in the literature for cost and performance for capture as a function of different CO2 concentrations. Energy requirements for physical and chemical solvents and solid looping are modelled as depending on $CO_2$ concentration as a function:

Power required ($GJ_{elec}/tCO_2$ captured) =

Reference value x adjustment factor,

Where adjustment factor = $11.22*[\%CO_2]^{-0.99}$

Steam required ($GJ_{heat}/tCO_2$ captured) =

Reference value x adjustment factor,

Where adjustment factor = $1.42*[\%CO_2]^{-0.142}$

These coefficients were derived from a curve-fit to data on MEA from Husebye *et al.* (2012, Upper left panel).

Given some fundamental similarities in the thermodynamic basis for the reactions, we assume this relationship holds broadly for 2nd generation chemical solvents, physical solvents and calcium looping, although no public reports are available, so this could benefit from experimental testing.

For cryogenics, power demand is modelled as depending on $CO_2$ concentration as follows:

Power required ($GJ/tCO_2$ captured) = $5.809*[\%CO_2]^{-0.657}$

This results from a curve-fit to Tunier *et al* (2011, bottom left panel).

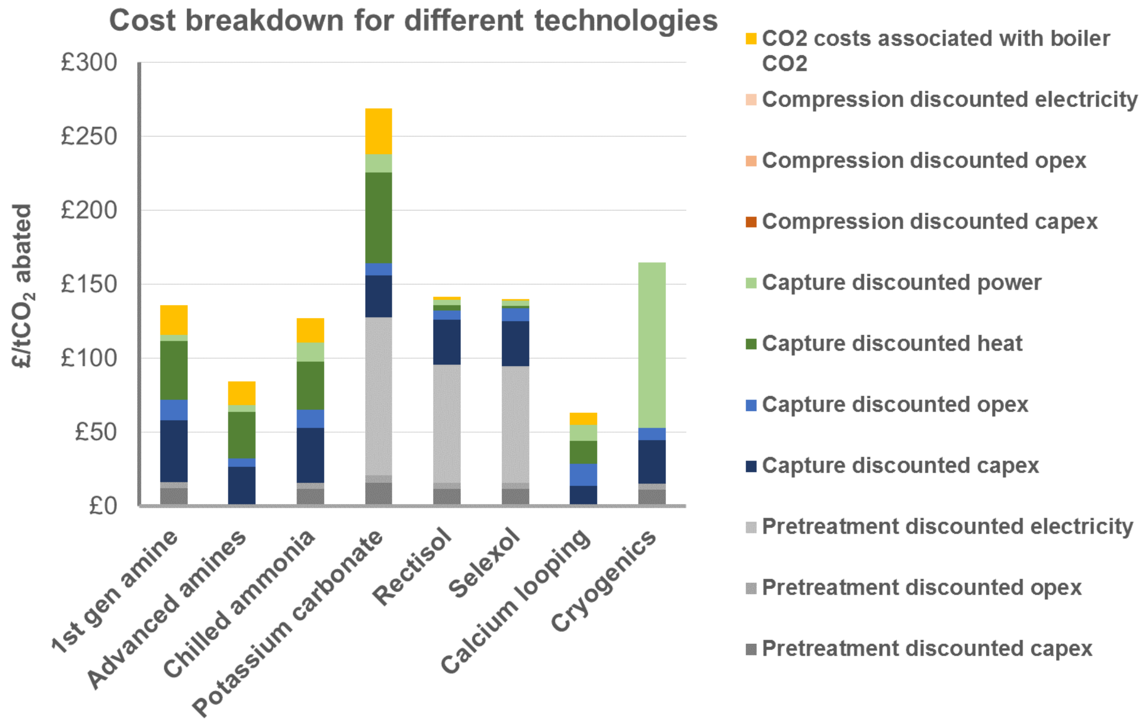
Husebye, J. *et al*. (2012) Techno-economic evaluation of amine-based CO2 capture: impact of CO2 concentration and steam supply *Energy Procedia* **23** 381-390; Tuinier *et al*. (2011)

# Techno-economic analysis

- Model Architecture
- Key assumptions
- Pragmatic scenario
- Sensitivity analysis

# The primary output from the techno-economic modelling is a high-level assessment of costs for different technologies.

| Input Parameter | Value |
|---|---|
| Source ID | #6 (Dunbar Cement) |
| Source size | 0.5 Mt/yr |
| Source purity | 24% $CO_2$ with 100 ppm NOx and 100 ppm SOx Flue gas stream at 1 bar. |
| Energy and carbon prices | DECC Central |
| Real discount rate | 10% |
| Lifetime | 15 yrs |
| Technology development scenario | Pragmatic |
| Site constraints | None |
| Timing | Construction 2022-2024 Yr of first operation 2025 |




Example illustrative industrial source for a range of capture technologies likely to be available in 2025.
Data shown is from a 1 bar, 0.5Mt$CO_2$/yr source with 24%$CO_2$, 0.44 Mt/yr captured, *ca*. 0.32-0.43 Mt$CO_2$/yr abated).

# Example model calculations – site $CO_2$ balance

| Name | Captured $CO_2$ emissions (tCO2/yr) | $CO_2$ from gas boiler (t/yr) | $CO_2$ associated with electricity (t/yr) | Abated $CO_2$ emissions (tCO2/yr) | Vented $CO_2$ emissions (tCO2/yr) |
|---|---|---|---|---|---|
| 1st gen amine | 437,481 | 80,747 | 2,081 | 354,654 | 134,266 |
| Advanced amines or blends | 437,481 | 67,289 | 2,081 | 368,112 | 120,808 |
| Chilled ammonia | 437,481 | 67,289 | 6,242 | 363,951 | 120,808 |
| Potassium carbonate | 437,481 | 112,148 | 5,201 | 320,132 | 165,667 |
| Rectisol | 437,481 | 8,972 | 2,081 | 426,429 | 62,491 |
| Selexol | 437,481 | 4,486 | 2,081 | 430,915 | 58,005 |
| Calcium looping | 413,177 | 33,894 | 5,305 | 373,978 | 111,717 |
| Cryogenics | 437,481 | - | 55,868 | 381,613 | 53,519 |

N.B. Vented emissions = initial source $CO_2$ – $CO_2$ captured +$CO_2$ from gas boiler (assumed not captured)

# Example model calculations – pre-treatment requirements

| Technology | Capex (£m) | Fixed opex (£m) | Heating (MWh) | Electricity (MWh) |
| --- | --- | --- | --- | --- |
| 1st gen amine | £ 29 | £ 1 | - | - |
| Advanced amines or blends | £ - | £ - | - | - |
| Chilled ammonia | £ 29 | £ 1 | - | - |
| Potassium carbonate | £ 35 | £ 2 | - | 251,475 |
| Rectisol | £ 35 | £ 2 | - | 251,475 |
| Selexol | £ 35 | £ 2 | - | 251,475 |
| Calcium looping | £ - | £ - | - | - |
| Cryogenics | £ 29 | £ 1 | - | - |

Costs exclude post-capture $CO_2$ compression, transport and storage.

# Example model calculations – capex and opex for capture

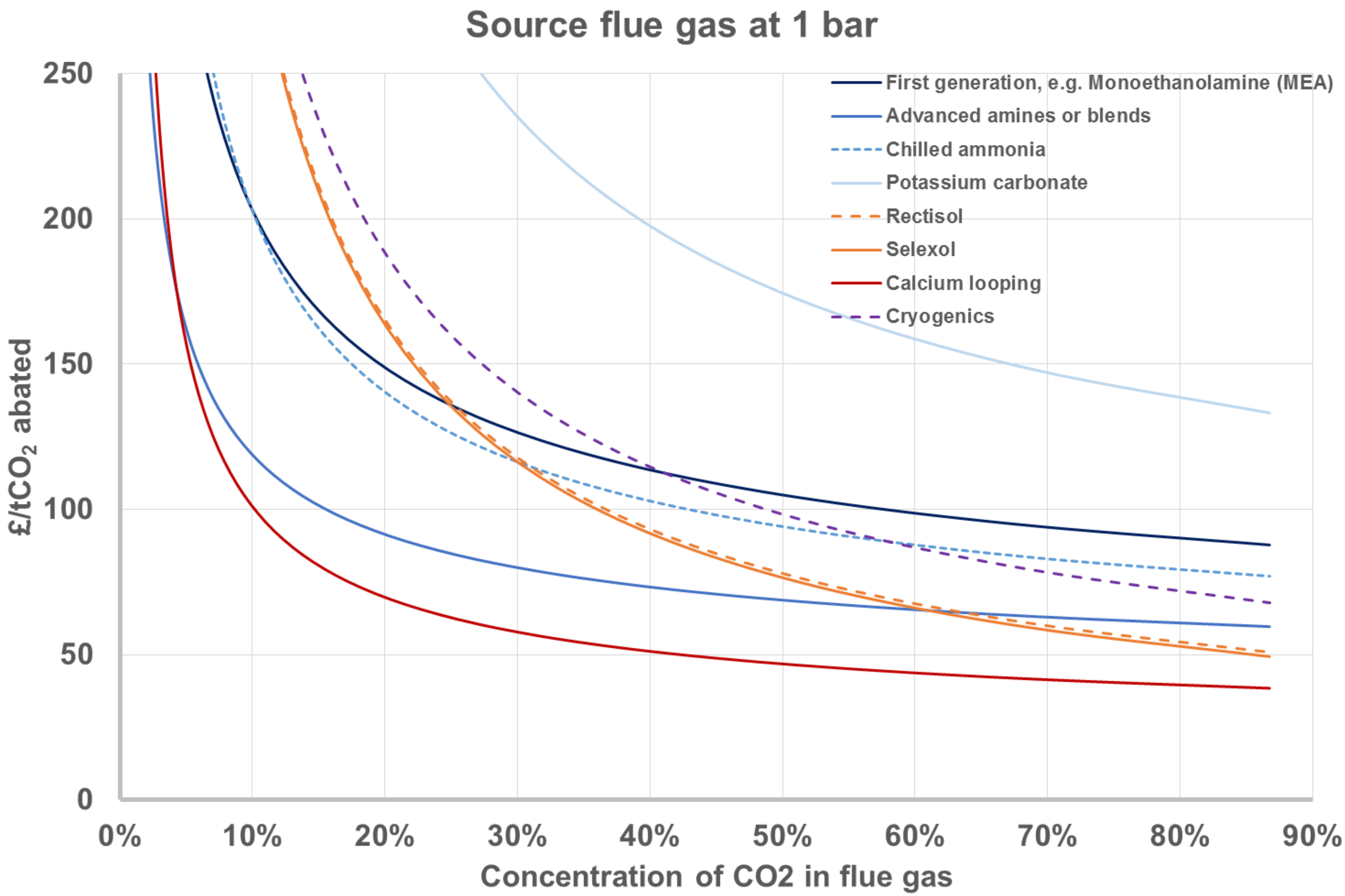
| Name | Capex (£m) | Fixed opex (£m/yr) | GJ heat/ $tCO_2$ captured | GJ electricity/ $tCO_2$ captured | Heating fuel (MWh gas/yr) | Electricity (MWh/yr) |
|---|---|---|---|---|---|---|
| 1st gen amine | £ 102 | £5 | 3.3 | 0.1 | 486,090 | 24,305 |
| Advanced amines or blends | £ 67 | £2 | 2.7 | 0.1 | 405,075 | 24,305 |
| Chilled ammonia | £ 93 | £4 | 2.7 | 0.3 | 405,075 | 72,914 |
| Potassium carbonate | £ 63 | £3 | 4.5 | 0.2 | 675,125 | 60,761 |
| Rectisol | £ 89 | £3 | 0.36 | 0.1 | 54,010 | 24,305 |
| Selexol | £ 90 | £4 | 0.18 | 0.1 | 27,005 | 24,305 |
| Calcium looping | £ 36 | £5 | 1.5 | 0.3 | 204,038 | 61,976 |
| Cryogenics | £ 78 | £3 | 0 | 2.6 | - | 437,481 |

# Example model outputs – overall capture performance

| Name | Discounted lifetime cost (£m) | Levelised cost of capture ($£/tCO_2$ captured) | Levelised cost of abatement ($£/t\ CO_2$ abated) |
|---|---|---|---|
| First generation amine | £ 410 | £ 112 | £ 138 |
| Advanced amines or blends | £ 265 | £ 72 | £ 86 |
| Chilled ammonia | £ 392 | £ 107 | £ 129 |
| Potassium carbonate | £ 729 | £ 199 | £ 272 |
| Rectisol | £ 506 | £ 138 | £ 142 |
| Selexol | £ 505 | £ 138 | £ 140 |
| Calcium looping | £ 200 | £ 58 | £ 64 |
| Cryogenics | £ 526 | £ 144 | £ 165 |

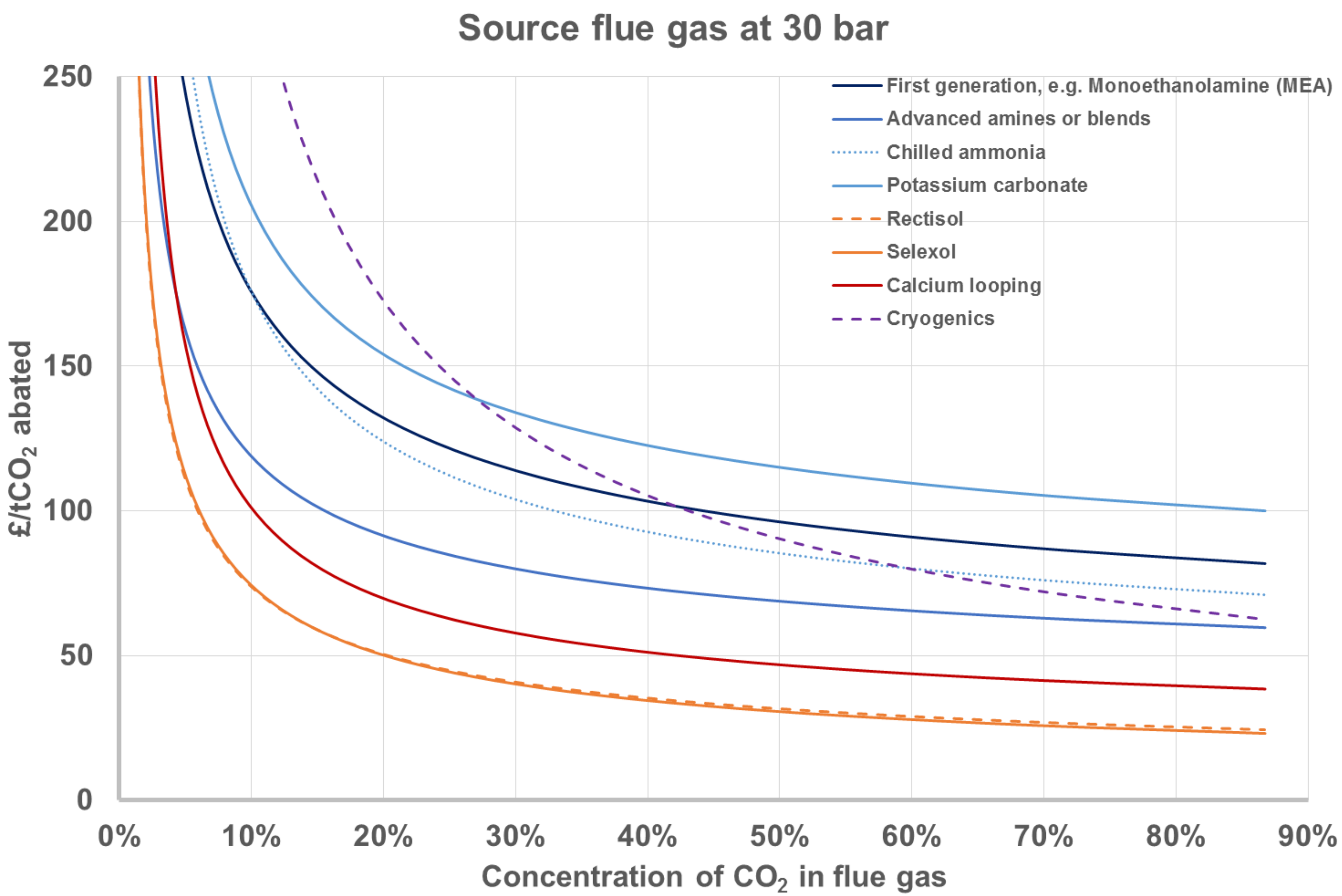
**N.B. Costs exclude compression, transport and storage**

# The model shows the that technology relevance is linked to $CO_2$ concentration and pressure (1/2) Low pressure



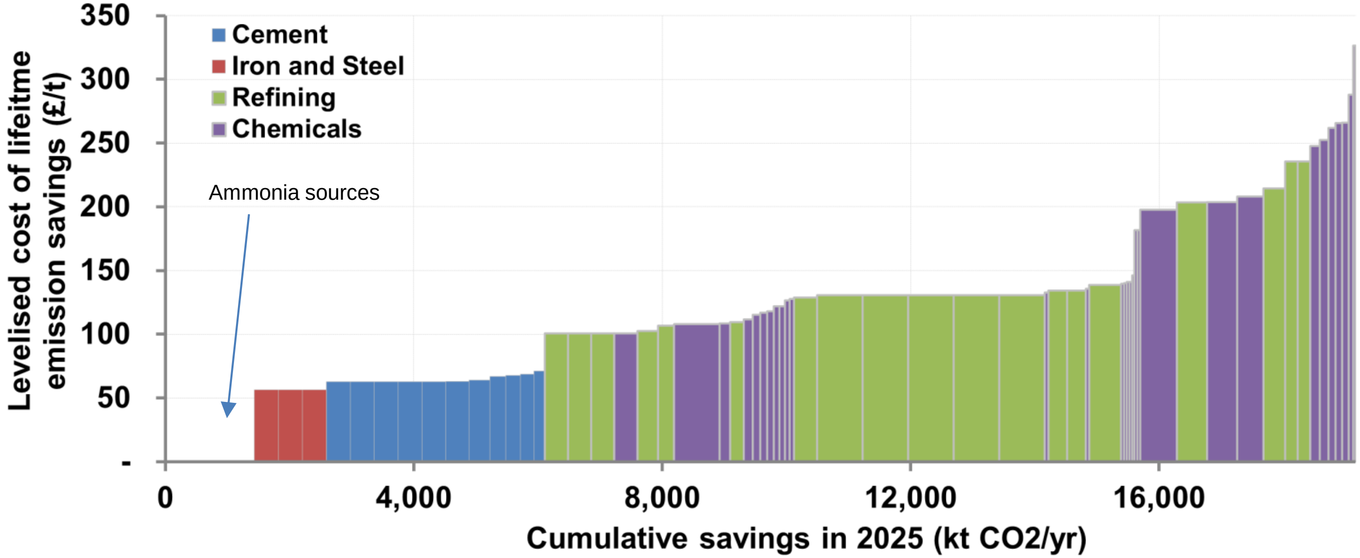

Other parameters as per 0.5 Mt/yr source in pragmatic scenario

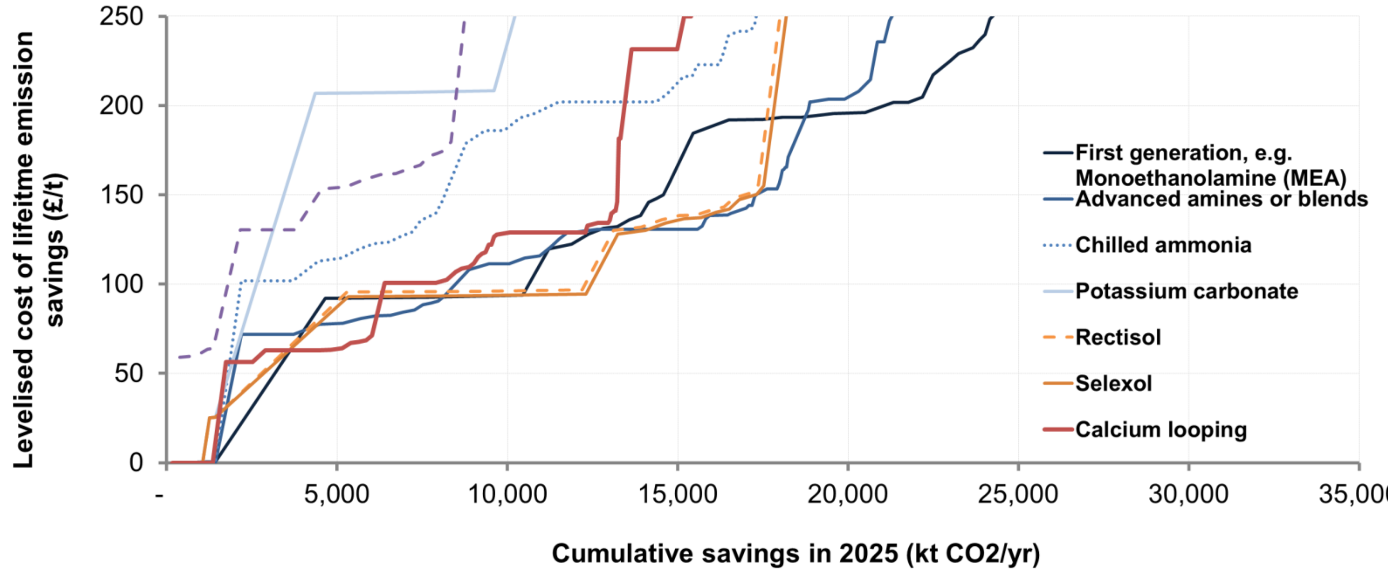
# The model shows the that technology relevance is linked to $CO_2$ concentration and pressure (2/2) High pressure




Other parameters as per 0.5 Mt/yr source in pragmatic scenario

# Pragmatic scenario – Comparison of MACCs for different technologies



# Assessment of technology-sector combinations

- Whilst the techno-economic model supports screening of capture technologies and sites, the differences between the outputs in terms of capture technologies or sectors are small relative to the model resolution.
- Matrices has been prepared that identifies attributes for each technology –source archetype contribution for a given scenario.
- High $CO_2$ purity hydrogen and ammonia sources have been excluded from this analysis as no capture technology is required.
- Attributes identified are
  - for projects: capex, fixed opex, gas requirement, electricity requirement, abatement cost, abatement potential, for chosen scenario
    - N.B. These are the individual median values of all relevant sector-technology combinations, i.e. median capex of all capexes, median opex of all opexes.
  - For sectors: max and min abatement cost, overall sector abatement potential in 2025 for chosen scenario
- These have been colour coded through a traffic light system as follows (green = favourable, red = unfavourable, yellow = intermediate).
- The following slides show results in the Pragmatic scenario.
- Given the uncertainties, readers should focus on trends rather than absolute numbers.

# Median capex, opex, gas and electricity requirements in the Pragmatic Scenario

| Median capture capex/£m | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Petrochemical cracker (olefins) | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £378 | £102 | £292 | £51 | £203 | £190 | £53 |
| 2nd gen chemical solvent | £97 | £67 | £173 | £34 | £134 | £125 | £35 |
| Chilled ammonia | £134 | £93 | £240 | £47 | £186 | £174 | £48 |
| Potassium carbonate | £233 | £63 | £180 | £31 | £125 | £117 | £33 |
| Rectisol | £328 | £89 | £253 | £44 | £176 | £165 | £46 |
| Selexol | £334 | £90 | £258 | £45 | £180 | £168 | £47 |
| Calcium looping | £32 | £36 | £58 | £18 | £71 | £58 | £19 |
| Cryogenics | £112 | £78 | £200 | £39 | £154 | £145 | £40 |

| Median capture fixed opex/£m/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £ 18 | £ 5 | £ 14 | £ 2 | £ 10 | £ 9 | £ 3 |
| 2nd gen chemical solvent | £ 3 | £ 2 | £ 5 | £ 1 | £ 4 | £ 4 | £ 1 |
| Chilled ammonia | £ 6 | £ 4 | £ 12 | £ 2 | £ 9 | £ 8 | £ 2 |
| Potassium carbonate | £ 10 | £ 3 | £ 8 | £ 1 | £ 5 | £ 5 | £ 1 |
| Rectisol | £ 10 | £ 3 | £ 8 | £ 1 | £ 5 | £ 5 | £ 1 |
| Selexol | £ 13 | £ 4 | £ 10 | £ 2 | £ 7 | £ 7 | £ 2 |
| Calcium looping | £ 5 | £ 5 | £ 9 | £ 3 | £ 11 | £ 9 | £ 3 |
| Cryogenics | £ 4 | £ 3 | £ 8 | £ 2 | £ 6 | £ 6 | £ 2 |

| Median gas requirement/TWh gas/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 3.62 | 0.44 | 1.20 | 0.07 | 0.32 | 0.63 | 0.10 |
| 2nd gen chemical solvent | 0.73 | 0.37 | 0.85 | 0.06 | 0.26 | 0.53 | 0.08 |
| Chilled ammonia | 0.73 | 0.37 | 0.85 | 0.06 | 0.26 | 0.53 | 0.08 |
| Potassium carbonate | 5.02 | 0.61 | 1.66 | 0.10 | 0.44 | 0.88 | 0.14 |
| Rectisol | 0.40 | 0.05 | 0.13 | 0.01 | 0.04 | 0.07 | 0.01 |
| Selexol | 0.20 | 0.02 | 0.07 | 0.00 | 0.02 | 0.04 | 0.01 |
| Calcium looping | 0.18 | 0.18 | 0.21 | 0.03 | 0.13 | 0.21 | 0.04 |
| Cryogenics | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |

| Median power requirement/TWhelec/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| 2nd gen chemical solvent | 0.05 | 0.02 | 0.05 | 0.00 | 0.01 | 0.03 | 0.00 |
| Chilled ammonia | 0.15 | 0.07 | 0.15 | 0.01 | 0.04 | 0.09 | 0.01 |
| Potassium carbonate | 0.52 | 0.06 | 0.15 | 0.01 | 0.03 | 0.08 | 0.01 |
| Rectisol | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Selexol | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Calcium looping | 0.06 | 0.06 | 0.06 | 0.01 | 0.03 | 0.06 | 0.01 |
| Cryogenics | 0.90 | 0.44 | 0.90 | 0.06 | 0.24 | 0.55 | 0.09 |

Costs exclude compression, transport and storage.

# Median abatement costs and potential in the Pragmatic Scenario

| | Pragmatic scenario | | | | | | |
|---|---|---|---|---|---|---|---|
| Median project abatement cost (£/tCO2 abated) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | £ 93 | £ 136 | £ 202 | £ 314 | £ 365 | £ 203 | £ 292 |
| 2nd gen chemical solvent | £ 72 | £ 84 | £ 131 | £ 202 | £ 242 | £ 115 | £ 150 |
| Chilled ammonia | £ 102 | £ 127 | £ 202 | £ 313 | £ 430 | £ 194 | £ 267 |
| Potassium carbonate | £ 207 | £ 269 | £ 536 | £ 844 | £ 1,747 | £ 531 | £ 571 |
| Rectisol | £ 96 | £ 142 | £ 303 | £ 513 | £ 950 | £ 302 | £ 345 |
| Selexol | £ 94 | £ 140 | £ 300 | £ 514 | £ 943 | £ 299 | £ 346 |
| Calcium looping | £ 56 | £ 63 | £ 129 | £ 182 | £ 256 | £ 101 | £ 127 |
| Cryogenics | £ 131 | £ 165 | £ 326 | £ 481 | £ 1,011 | £ 315 | £ 363 |
| | | | | | | | |
| Median individual project abatement potential (MtCO2/yr in 2025) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | 3.0 | 0.4 | 0.8 | 0.04 | 0.2 | 0.4 | 0.1 |
| 2nd gen chemical solvent | 0.8 | 0.4 | 0.7 | 0.05 | 0.2 | 0.5 | 0.1 |
| Chilled ammonia | 0.8 | 0.4 | 0.7 | 0.04 | 0.2 | 0.4 | 0.1 |
| Potassium carbonate | 2.8 | 0.3 | 0.7 | 0.04 | 0.1 | 0.4 | 0.1 |
| Rectisol | 3.6 | 0.4 | 1.0 | 0.06 | 0.2 | 0.5 | 0.1 |
| Selexol | 3.7 | 0.4 | 1.0 | 0.06 | 0.2 | 0.5 | 0.1 |
| Calcium looping | 0.4 | 0.4 | 0.4 | 0.05 | 0.2 | 0.4 | 0.1 |
| Cryogenics | 0.8 | 0.4 | 0.7 | 0.04 | 0.1 | 0.4 | 0.1 |
| | | | | | | | |
| Median 2025 project discounted lifetime cost (£m, 10%, 15 yrs) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | £2,352 | £404 | £1,383 | £116 | £512 | £736 | £167 |
| 2nd gen chemical solvent | £458 | £259 | £802 | £78 | £358 | £436 | £90 |
| Chilled ammonia | £644 | £386 | £1,206 | £116 | £575 | £714 | £156 |
| Potassium carbonate | £4,783 | £720 | £3,209 | £268 | £1,931 | £1,678 | £287 |
| Rectisol | £2,913 | £505 | £2,574 | £239 | £1,746 | £1,351 | £244 |
| Selexol | £2,870 | £505 | £2,576 | £242 | £1,759 | £1,356 | £247 |
| Calcium looping | £183 | £197 | £402 | £71 | £374 | £314 | £77 |
| Cryogenics | £875 | £526 | £1,895 | £166 | £992 | £1,130 | £209 |

Costs exclude compression, transport and storage.

# Overall sectoral attributes in the Pragmatic Scenario

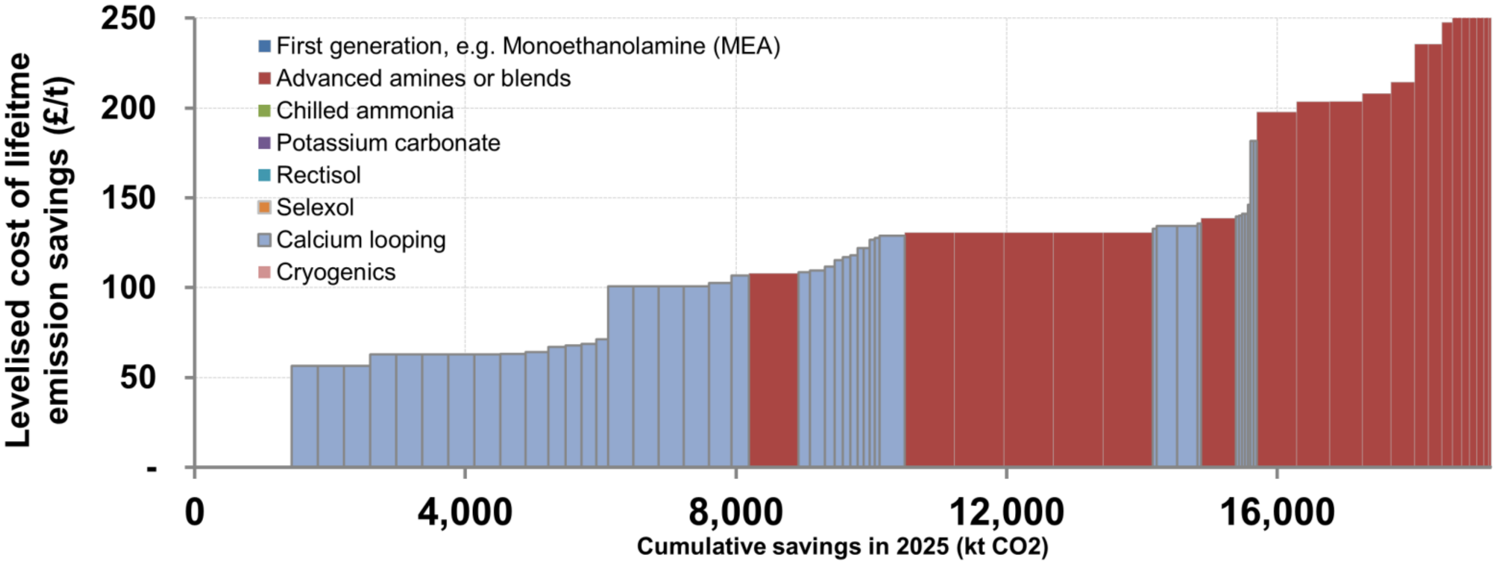
| Combined abatement potential (Mt/yr) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 9.0 | 4.3 | 6.2 | 0.1 | 3.3 | 3.5 | 1.2 |
| 2nd gen chemical solvent | 2.3 | 4.4 | 5.2 | 0.1 | 3.5 | 3.6 | 1.3 |
| Chilled ammonia | 2.3 | 4.4 | 5.1 | 0.1 | 3.1 | 3.5 | 1.2 |
| Potassium carbonate | 8.2 | 3.9 | 5.4 | 0.1 | 2.6 | 3.0 | 1.1 |
| Rectisol | 10.8 | 5.2 | 7.7 | 0.1 | 4.3 | 4.3 | 1.5 |
| Selexol | 10.9 | 5.2 | 7.8 | 0.1 | 4.4 | 4.3 | 1.5 |
| Calcium looping | 1.2 | 3.5 | 3.2 | 0.1 | 2.8 | 2.7 | 1.3 |
| Cryogenics | 2.4 | 4.6 | 4.9 | 0.1 | 2.3 | 3.4 | 1.2 |

| Range of pragmatic scenario abatement costs | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 92 - 94 | 120 - 157 | 192 - 254 | 314 - 314 | 288 - 516 | 184 - 237 | 240 - 346 |
| 2nd gen chemical solvent | 72 - 72 | 77 - 93 | 131 - 153 | 202 - 202 | 198 - 327 | 108 - 129 | 129 - 171 |
| Chilled ammonia | 102 - 102 | 113 - 146 | 202 - 242 | 313 - 313 | 355 - 574 | 179 - 223 | 223 - 313 |
| Potassium carbonate | 207 - 208 | 256 - 284 | 527 - 583 | 844 - 844 | 1644 - 1940 | 515 - 559 | 528 - 615 |
| Rectisol | 95 - 97 | 130 - 156 | 297 - 340 | 513 - 513 | 877 - 1088 | 288 - 325 | 309 - 382 |
| Selexol | 93 - 95 | 128 - 155 | 293 - 338 | 514 - 514 | 868 - 1086 | 285 - 323 | 308 - 384 |
| Calcium looping | 56 - 56 | 63 - 71 | 129 - 134 | 182 - 182 | 232 - 333 | 101 - 110 | 109 - 146 |
| Cryogenics | 131 - 131 | 153 - 180 | 326 - 361 | 481 - 481 | 930 - 1168 | 302 - 340 | 325 - 402 |

| Number of projects | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| 2nd gen chemical solvent | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Chilled ammonia | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Potassium carbonate | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Rectisol | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Selexol | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Calcium looping | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Cryogenics | 3 | 11 | 9 | 2 | 14 | 8 | 16 |

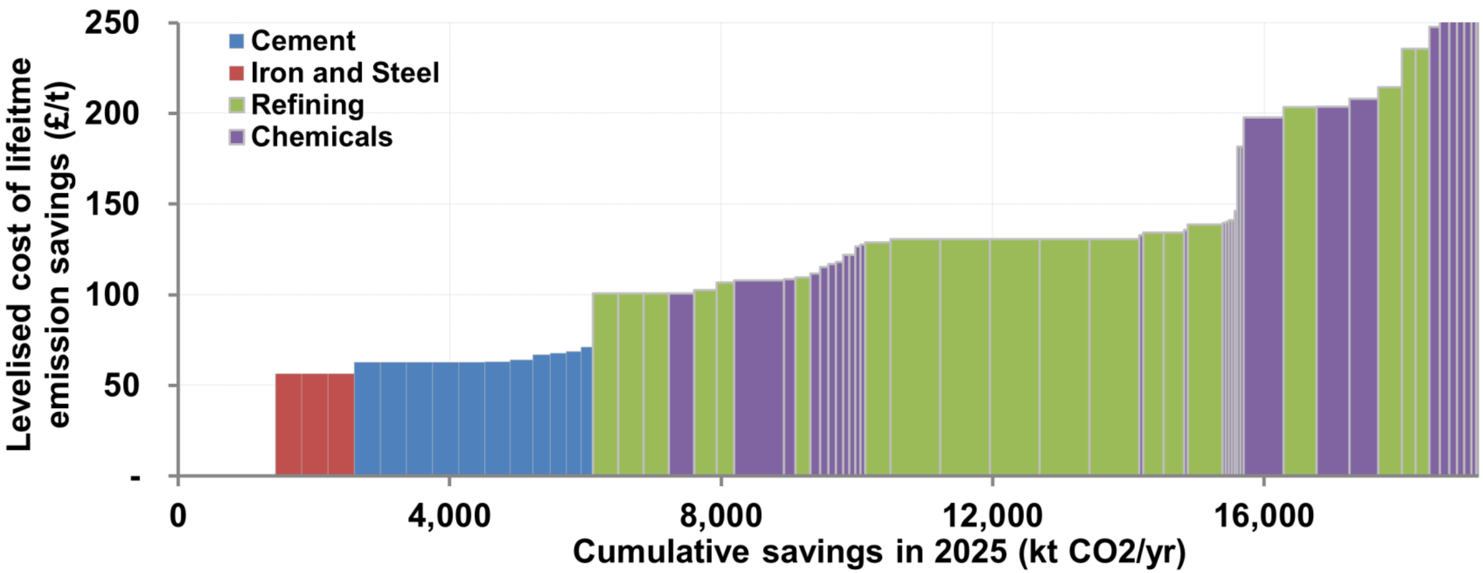
Costs exclude compression, transport and storage.

# Comparison of the least cost (£/t$CO_2$ abated) technologies and sectors in the pragmatic scenario




**Technologies:**

- calcium looping and
- $2^{nd}$ generation chemical solvents



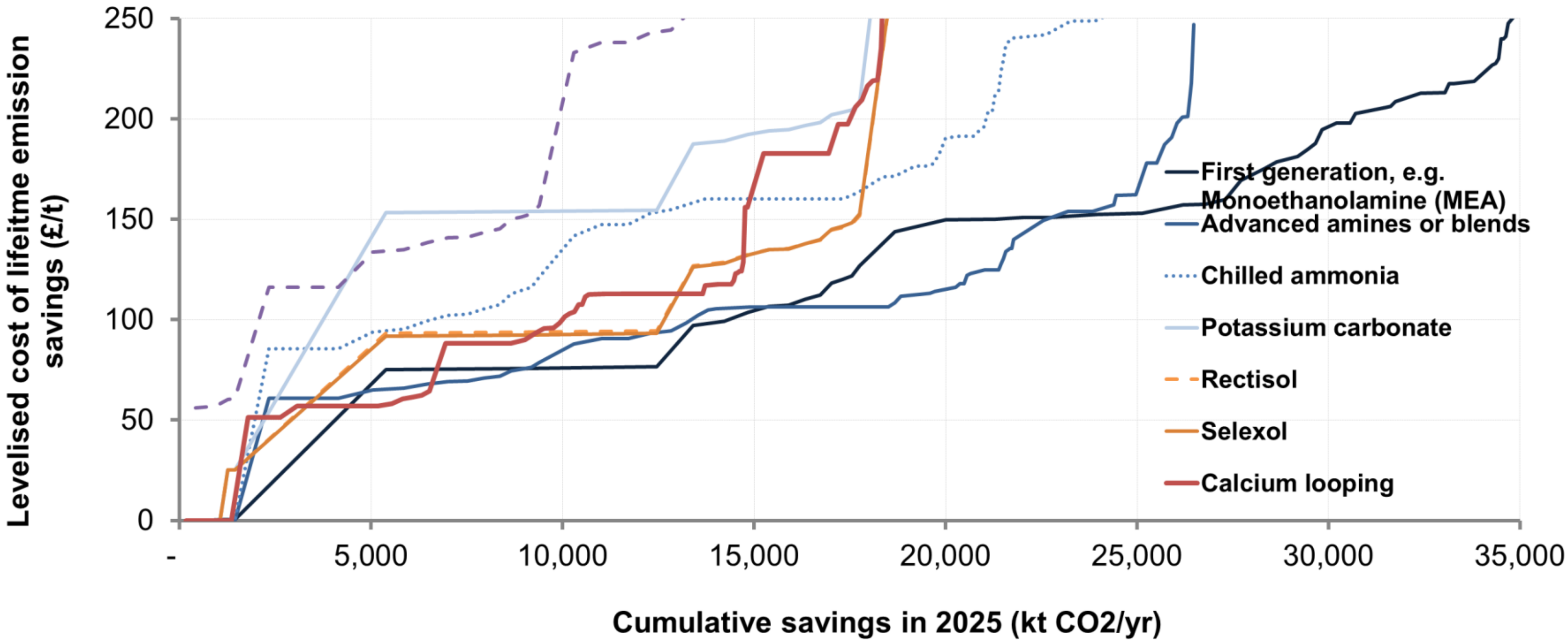

**Sectors:**

- Iron and Steel
- Cement

(high purity sectors do not require additional separation)

# Heat and power consumption energy imply significant differences between $CO_2$ captured and $CO_2$ abated.

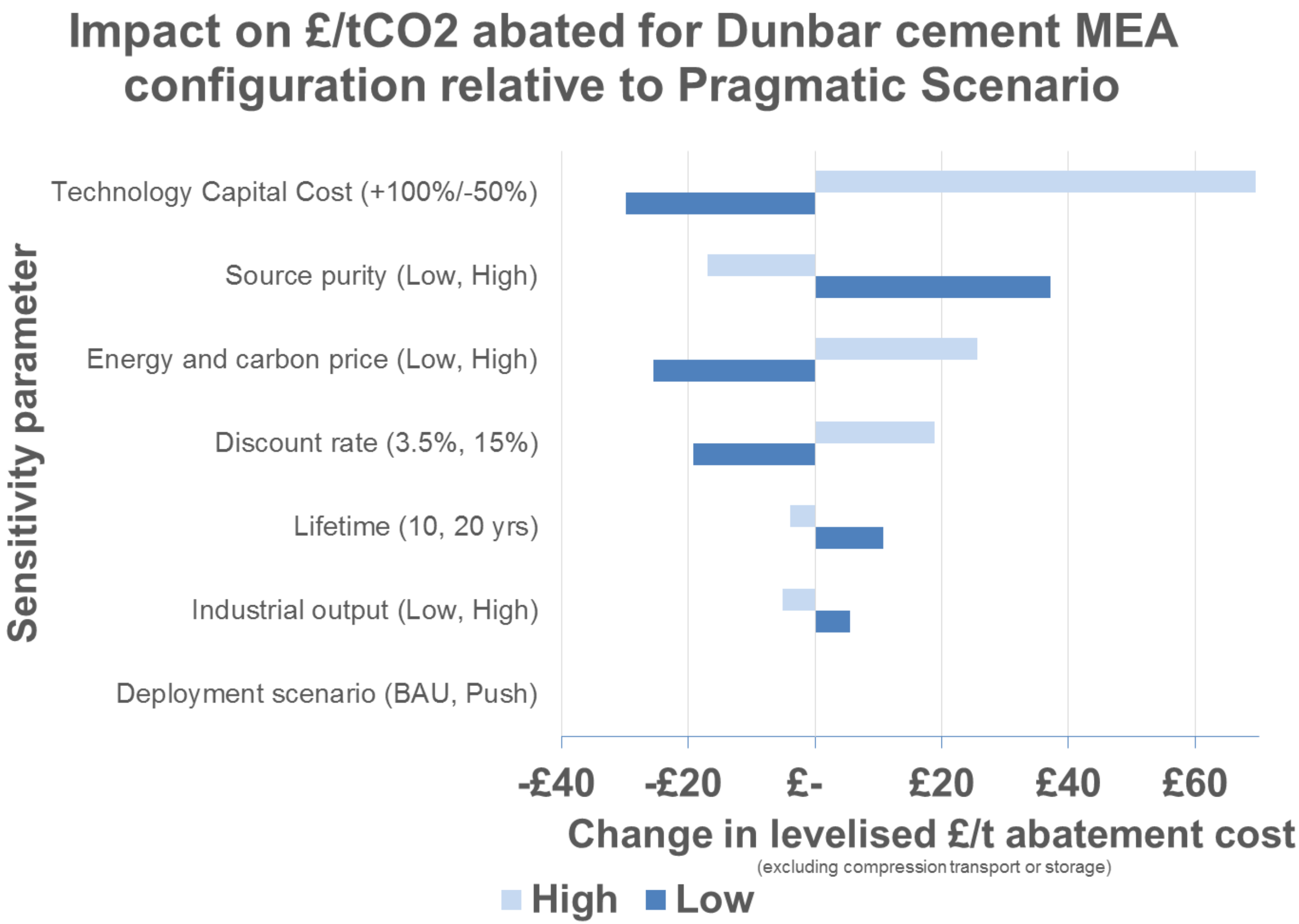
£/t$CO_2$ captured in the Pragmatic scenario



# Techno-economic analysis

- Model Architecture
- Key assumptions
- Pragmatic scenario
- Sensitivity analysis

# The techno-economic model facilitates high level sensitivity analysis for a source-technology combination, allowing uncertainties to be prioritised.



# Illustrative technology readiness level milestones for industrial $CO_2$ capture at sources of 0.1M-1M tCO2/yr

**TRL 3**
Detailed engineering studies & simulations
Bench scale prototypes e.g. $10^1 tCO_2$/yr

**TRL 4**
Lab scale testing in operation e.g. 100 $tCO_2$/yr
Pilots planned

**TRL 5**
Lab scale successful *ca*. 1kt$CO_2$/yr
Pilot plants in development

**TRL 6**
Pilot plant successful at operation *ca*. 10kt$CO_2$/yr
Small demo in development

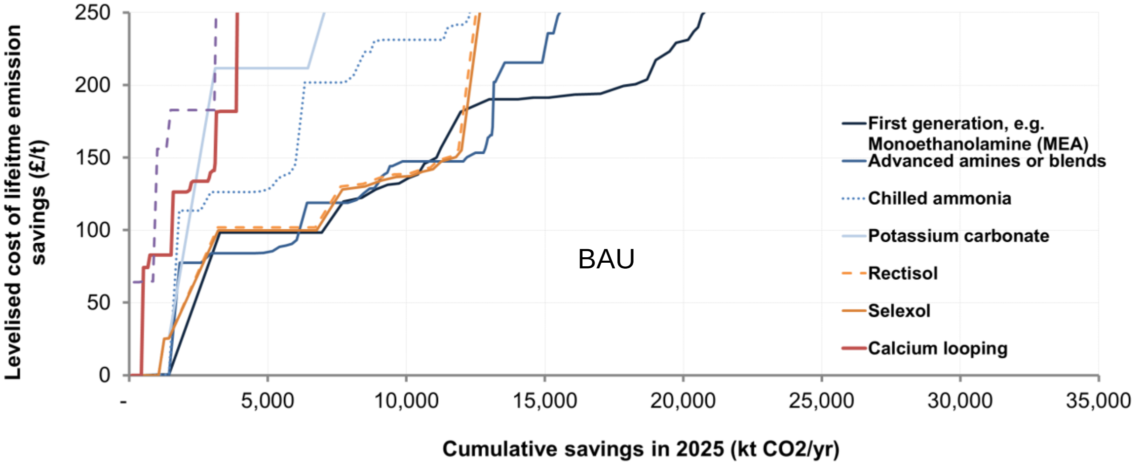
**TRL 7**
1st Small demonstration successful in operation *ca*. 0.1M $tCO_2$/yr
Large demo in development

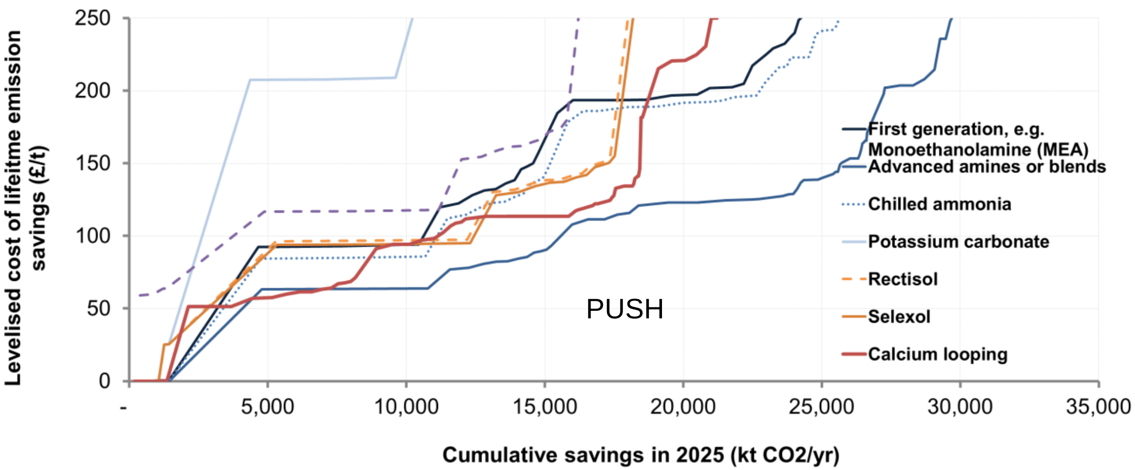
**TRL 8**
Demonstration *ca*. 0.1-1Mt$CO_2$/yr successful, potential for optimisation.
Commercial scale plant in development.

**TRL 9**
Commercially mature, with systems and components successfully optimised for operation at *ca*.1Mt$CO_2$/yr

- TRL classification is approximate (and the same technology can be at different levels for different applications)
- With no published TRL assignations for industrial capture, TRLs have been estimated for this study from published descriptions of pilots/demos against an eventual requirement to meet UK industrial $CO_2$ capture scale requirements.
- Estimates of scale corresponding to TRL are illustrative only.
- Business As Usual, e.g. *one* TRL increase or <1 order of magnitude increase in capacity from today for 2020-2025
- Pragmatic Scenario, e.g. *two* TRL increases or *ca.* 1 order of magnitude increase in capacity from today for 2020-2025
- Push Scenario, e.g. up to *three* TRL increases or *ca*. 2 orders of magnitude increase in capacity from today for 2020-2025

# Sensitivity of Technology MACCs to Technology Deployment Rate Scenario

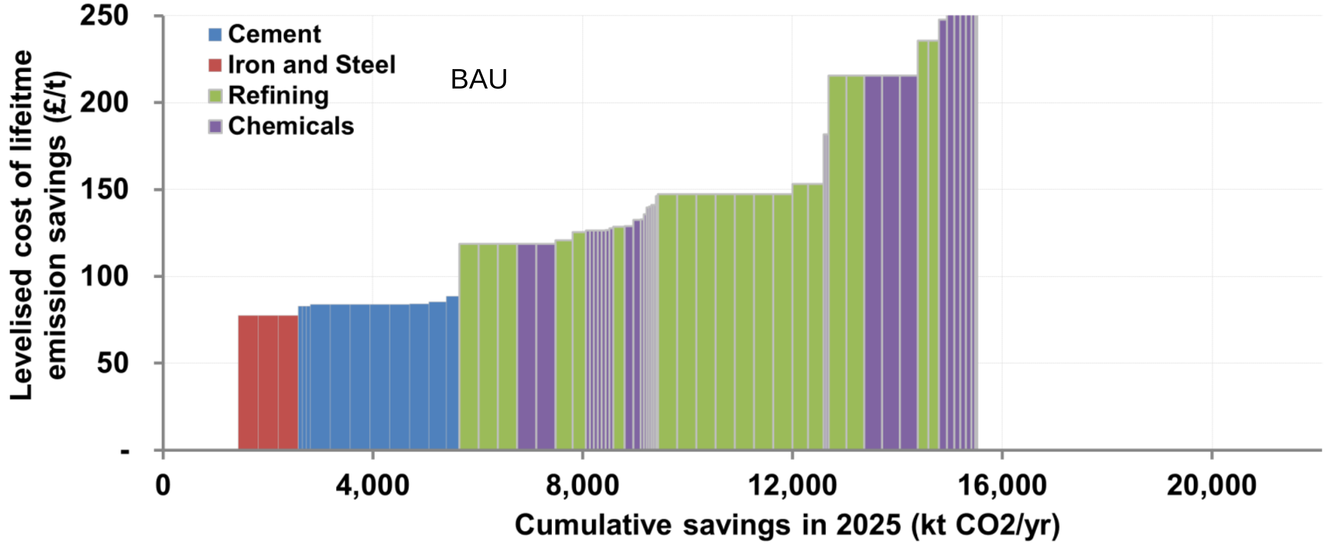


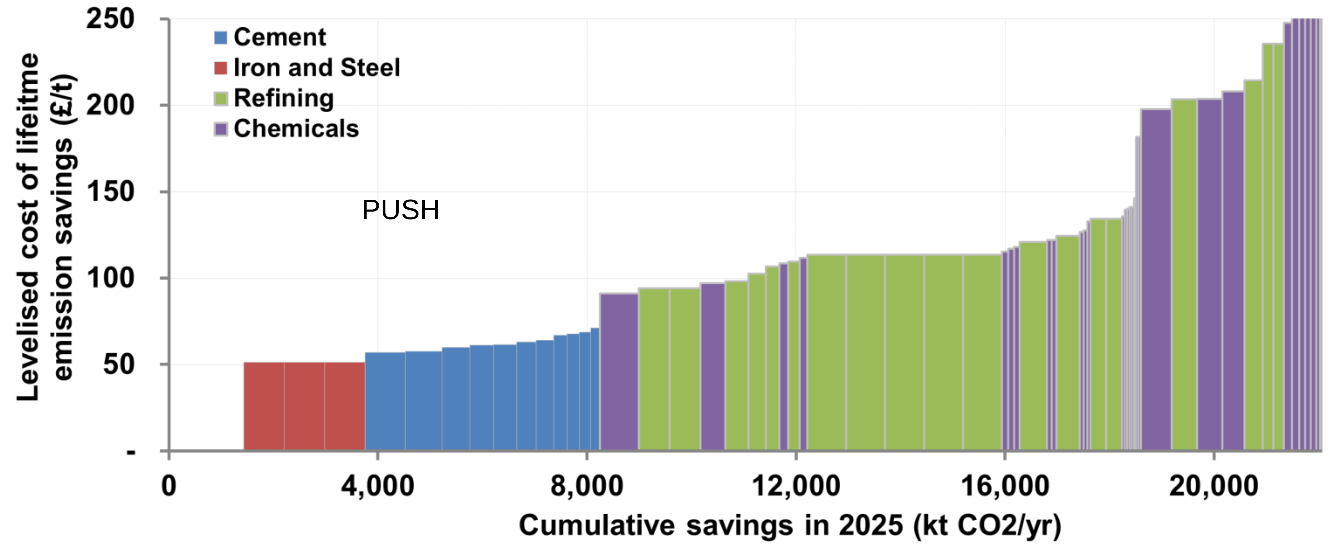




| | Maximum scale of deployment (Mt/yr) in 2025 | | |
|---|---|---|---|
| Technology | BAU | Pragmatic | Push |
| 1st gen amine | 2.5 | 5 | 10 |
| 2nd gen amine | 0.5 | 1 | 5 |
| Ammonia | 0.5 | 1 | 5 |
| Potassium carbonate | 2.5 | 5 | 10 |
| Rectisol | 2 | 5 | 10 |
| Selexol | 2 | 5 | 10 |
| Calcium looping | 0.1 | 0.5 | 1 |
| Cryogenics | 0.2 | 1 | 5 |

Other conditions as per Pragmatic Scenario. Costs exclude compression, transport and storage.

# Sensitivity of sector marginal abatement costs and potentials to technology deployment rate scenario.




Other conditions as per Pragmatic Scenario. Costs exclude compression, transport and storage.

# Median project capex, opex, gas and electricity requirements for BAU scenario

| Median capture capex/£m | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £271 | £102 | £292 | £51 | £203 | £190 | £53 |
| 2nd gen chemical solvent | £61 | £67 | £109 | £34 | £134 | £109 | £35 |
| Chilled ammonia | £85 | £93 | £151 | £47 | £186 | £151 | £48 |
| Potassium carbonate | £167 | £63 | £180 | £31 | £125 | £117 | £33 |
| Rectisol | £202 | £89 | £253 | £44 | £176 | £165 | £46 |
| Selexol | £206 | £90 | £258 | £45 | £180 | £168 | £47 |
| Calcium looping | £11 | £12 | £20 | £18 | £38 | £20 | £19 |
| Cryogenics | £38 | £43 | £68 | £39 | £129 | £68 | £40 |

| Median capture fixed opex/£m/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £ 13 | £ 5 | £ 14 | £ 2 | £ 10 | £ 9 | £ 3 |
| 2nd gen chemical solvent | £ 2 | £ 2 | £ 3 | £ 1 | £ 4 | £ 3 | £ 1 |
| Chilled ammonia | £ 4 | £ 4 | £ 7 | £ 2 | £ 9 | £ 7 | £ 2 |
| Potassium carbonate | £ 7 | £ 3 | £ 8 | £ 1 | £ 5 | £ 5 | £ 1 |
| Rectisol | £ 6 | £ 3 | £ 8 | £ 1 | £ 5 | £ 5 | £ 1 |
| Selexol | £ 8 | £ 4 | £ 10 | £ 2 | £ 7 | £ 7 | £ 2 |
| Calcium looping | £ 2 | £ 2 | £ 3 | £ 3 | £ 6 | £ 3 | £ 3 |
| Cryogenics | £ 2 | £ 2 | £ 3 | £ 2 | £ 5 | £ 3 | £ 2 |

| Median gas requirement/TWh gas/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 2.19 | 0.44 | 1.20 | 0.07 | 0.32 | 0.63 | 0.10 |
| 2nd gen chemical solvent | 0.37 | 0.37 | 0.43 | 0.06 | 0.26 | 0.43 | 0.08 |
| Chilled ammonia | 0.37 | 0.37 | 0.43 | 0.06 | 0.26 | 0.43 | 0.08 |
| Potassium carbonate | 3.04 | 0.61 | 1.66 | 0.10 | 0.44 | 0.88 | 0.14 |
| Rectisol | 0.19 | 0.05 | 0.13 | 0.01 | 0.04 | 0.07 | 0.01 |
| Selexol | 0.10 | 0.02 | 0.07 | 0.00 | 0.02 | 0.04 | 0.01 |
| Calcium looping | 0.04 | 0.04 | 0.04 | 0.03 | 0.05 | 0.04 | 0.04 |
| Cryogenics | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |

| Median power requirement/TWhelec/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 0.13 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| 2nd gen chemical solvent | 0.03 | 0.02 | 0.03 | 0.00 | 0.01 | 0.03 | 0.00 |
| Chilled ammonia | 0.08 | 0.07 | 0.08 | 0.01 | 0.04 | 0.08 | 0.01 |
| Potassium carbonate | 0.31 | 0.06 | 0.15 | 0.01 | 0.03 | 0.08 | 0.01 |
| Rectisol | 0.10 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Selexol | 0.10 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Calcium looping | 0.01 | 0.01 | 0.01 | 0.01 | 0.01 | 0.01 | 0.01 |
| Cryogenics | 0.18 | 0.18 | 0.18 | 0.06 | 0.18 | 0.18 | 0.09 |

Other conditions as per Pragmatic Scenario

# Median Project Abatement Costs and Potential in the BAU Scenario

| | Business As Usual | | | | | | |
|---|---|---|---|---|---|---|---|
| Median project abatement cost (£/tCO2 abated) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | £ 98 | £ 136 | £ 199 | £ 314 | £ 365 | £ 202 | £ 292 |
| 2nd gen chemical solvent | £ 78 | £ 84 | £ 147 | £ 202 | £ 242 | £ 119 | £ 150 |
| Chilled ammonia | £ 113 | £ 127 | £ 231 | £ 313 | £ 430 | £ 202 | £ 267 |
| Potassium carbonate | £ 212 | £ 269 | £ 533 | £ 844 | £ 1,747 | £ 529 | £ 571 |
| Rectisol | £ 102 | £ 142 | £ 301 | £ 513 | £ 950 | £ 301 | £ 345 |
| Selexol | £ 100 | £ 140 | £ 297 | £ 514 | £ 943 | £ 298 | £ 346 |
| Calcium looping | £ 74 | £ 83 | £ 182 | £ 182 | £ 301 | £ 134 | £ 127 |
| Cryogenics | £ 156 | £ 183 | £ 397 | £ 481 | £ 1,033 | £ 356 | £ 363 |

| Median individual project abatement potential (MtCO2/yr in 2025) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 1.8 | 0.4 | 0.8 | 0.04 | 0.2 | 0.4 | 0.1 |
| 2nd gen chemical solvent | 0.4 | 0.4 | 0.4 | 0.05 | 0.2 | 0.4 | 0.1 |
| Chilled ammonia | 0.4 | 0.4 | 0.4 | 0.04 | 0.2 | 0.4 | 0.1 |
| Potassium carbonate | 1.7 | 0.3 | 0.7 | 0.04 | 0.1 | 0.4 | 0.1 |
| Rectisol | 1.8 | 0.4 | 1.0 | 0.06 | 0.2 | 0.5 | 0.1 |
| Selexol | 1.8 | 0.4 | 1.0 | 0.06 | 0.2 | 0.5 | 0.1 |
| Calcium looping | 0.1 | 0.1 | 0.1 | 0.05 | 0.1 | 0.1 | 0.1 |
| Cryogenics | 0.2 | 0.2 | 0.1 | 0.04 | 0.1 | 0.1 | 0.1 |

| Median 2025 project discounted lifetime cost (£m, 10%, 15 yrs) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £1,511 | £404 | £1,366 | £116 | £512 | £731 | £167 |
| 2nd gen chemical solvent | £247 | £259 | £452 | £78 | £358 | £364 | £90 |
| Chilled ammonia | £359 | £386 | £690 | £116 | £575 | £602 | £156 |
| Potassium carbonate | £2,958 | £720 | £3,192 | £268 | £1,931 | £1,674 | £287 |
| Rectisol | £1,497 | £505 | £2,552 | £239 | £1,746 | £1,347 | £244 |
| Selexol | £1,481 | £505 | £2,554 | £242 | £1,759 | £1,352 | £247 |
| Calcium looping | £48 | £53 | £113 | £71 | £169 | £83 | £77 |
| Cryogenics | £209 | £240 | £462 | £166 | £777 | £415 | £209 |

Other conditions as per Pragmatic Scenario

# Overall sectoral attributes in the BAU Scenario

| Combined abatement potential (Mt/yr) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 5.5 | 4.3 | 6.2 | 0.1 | 3.3 | 3.5 | 1.2 |
| 2nd gen chemical solvent | 1.1 | 3.5 | 3.2 | 0.1 | 2.8 | 2.6 | 1.3 |
| Chilled ammonia | 1.1 | 3.4 | 3.1 | 0.1 | 2.6 | 2.6 | 1.2 |
| Potassium carbonate | 5.0 | 3.9 | 5.4 | 0.1 | 2.6 | 3.0 | 1.1 |
| Rectisol | 5.3 | 5.2 | 7.7 | 0.1 | 4.3 | 4.3 | 1.5 |
| Selexol | 5.3 | 5.2 | 7.8 | 0.1 | 4.4 | 4.3 | 1.5 |
| Calcium looping | 0.2 | 0.8 | 0.7 | 0.1 | 0.9 | 0.6 | 1.0 |
| Cryogenics | 0.5 | 1.7 | 1.3 | 0.1 | 1.1 | 1.1 | 1.2 |

| Range of pragmatic scenario abatement costs | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 98 - 98 | 120 - 157 | 190 - 254 | 314 - 314 | 288 - 516 | 182 - 237 | 240 - 346 |
| 2nd gen chemical solvent | 78 - 78 | 84 - 93 | 147 - 153 | 202 - 202 | 215 - 327 | 119 - 129 | 129 - 171 |
| Chilled ammonia | 113 - 113 | 126 - 146 | 231 - 242 | 313 - 313 | 385 - 574 | 202 - 223 | 223 - 313 |
| Potassium carbonate | 212 - 212 | 256 - 284 | 524 - 583 | 844 - 844 | 1644 - 1940 | 511 - 559 | 528 - 615 |
| Rectisol | 102 - 102 | 130 - 156 | 294 - 340 | 513 - 513 | 877 - 1088 | 286 - 325 | 309 - 382 |
| Selexol | 100 - 100 | 128 - 155 | 290 - 338 | 514 - 514 | 868 - 1086 | 283 - 323 | 308 - 384 |
| Calcium looping | 74 - 74 | 83 - 83 | 182 - 182 | 182 - 182 | 301 - 333 | 134 - 134 | 126 - 146 |
| Cryogenics | 156 - 156 | 183 - 183 | 397 - 397 | 481 - 481 | 1033 - 1168 | 356 - 356 | 330 - 402 |

| Number of projects | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| 2nd gen chemical solvent | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Chilled ammonia | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Potassium carbonate | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Rectisol | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Selexol | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Calcium looping | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Cryogenics | 3 | 11 | 9 | 2 | 14 | 8 | 16 |

Other conditions as per Pragmatic Scenario

# Median project capex, opex, gas and electricity requirements in the Push Scenario

| Median capture capex/£m | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £378 | £102 | £292 | £51 | £203 | £190 | £53 |
| 2nd gen chemical solvent | £249 | £67 | £192 | £34 | £134 | £125 | £35 |
| Chilled ammonia | £345 | £93 | £266 | £47 | £186 | £174 | £48 |
| Potassium carbonate | £233 | £63 | £180 | £31 | £125 | £117 | £33 |
| Rectisol | £328 | £89 | £253 | £44 | £176 | £165 | £46 |
| Selexol | £334 | £90 | £258 | £45 | £180 | £168 | £47 |
| Calcium looping | £52 | £36 | £92 | £18 | £71 | £67 | £19 |
| Cryogenics | £287 | £78 | £221 | £39 | £154 | £145 | £40 |

| Median capture fixed opex/£m/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | £ 18 | £ 5 | £ 14 | £ 2 | £ 10 | £ 9 | £ 3 |
| 2nd gen chemical solvent | £ 7 | £ 2 | £ 6 | £ 1 | £ 4 | £ 4 | £ 1 |
| Chilled ammonia | £ 17 | £ 4 | £ 13 | £ 2 | £ 9 | £ 8 | £ 2 |
| Potassium carbonate | £ 10 | £ 3 | £ 8 | £ 1 | £ 5 | £ 5 | £ 1 |
| Rectisol | £ 10 | £ 3 | £ 8 | £ 1 | £ 5 | £ 5 | £ 1 |
| Selexol | £ 13 | £ 4 | £ 10 | £ 2 | £ 7 | £ 7 | £ 2 |
| Calcium looping | £ 8 | £ 5 | £ 14 | £ 3 | £ 11 | £ 10 | £ 3 |
| Cryogenics | £ 11 | £ 3 | £ 9 | £ 2 | £ 6 | £ 6 | £ 2 |

| Median gas requirement/TWh gas/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 3.62 | 0.44 | 1.20 | 0.07 | 0.32 | 0.63 | 0.10 |
| 2nd gen chemical solvent | 3.01 | 0.37 | 1.00 | 0.06 | 0.26 | 0.53 | 0.08 |
| Chilled ammonia | 3.01 | 0.37 | 1.00 | 0.06 | 0.26 | 0.53 | 0.08 |
| Potassium carbonate | 5.02 | 0.61 | 1.66 | 0.10 | 0.44 | 0.88 | 0.14 |
| Rectisol | 0.40 | 0.05 | 0.13 | 0.01 | 0.04 | 0.07 | 0.01 |
| Selexol | 0.20 | 0.02 | 0.07 | 0.00 | 0.02 | 0.04 | 0.01 |
| Calcium looping | 0.37 | 0.18 | 0.43 | 0.03 | 0.13 | 0.26 | 0.04 |
| Cryogenics | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |

| Median power requirement/TWhelec/yr | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| 2nd gen chemical solvent | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Chilled ammonia | 0.62 | 0.07 | 0.18 | 0.01 | 0.04 | 0.09 | 0.01 |
| Potassium carbonate | 0.52 | 0.06 | 0.15 | 0.01 | 0.03 | 0.08 | 0.01 |
| Rectisol | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Selexol | 0.21 | 0.02 | 0.06 | 0.00 | 0.01 | 0.03 | 0.00 |
| Calcium looping | 0.13 | 0.06 | 0.13 | 0.01 | 0.03 | 0.08 | 0.01 |
| Cryogenics | 3.72 | 0.44 | 1.05 | 0.06 | 0.24 | 0.55 | 0.09 |

Other conditions as per Pragmatic Scenario

# Median project abatement costs and potential in the Push scenario

| | Push scenario | | | | | | |
|---|---|---|---|---|---|---|---|
| Median project abatement cost (£/tCO2 abated) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | £ 93 | £ 136 | £ 203 | £ 314 | £ 365 | £ 203 | £ 292 |
| 2nd gen chemical solvent | £ 63 | £ 84 | £ 127 | £ 202 | £ 242 | £ 115 | £ 150 |
| Chilled ammonia | £ 85 | £ 127 | £ 197 | £ 313 | £ 430 | £ 194 | £ 267 |
| Potassium carbonate | £ 208 | £ 269 | £ 537 | £ 844 | £ 1,747 | £ 531 | £ 571 |
| Rectisol | £ 97 | £ 142 | £ 303 | £ 513 | £ 950 | £ 302 | £ 345 |
| Selexol | £ 94 | £ 140 | £ 300 | £ 514 | £ 943 | £ 299 | £ 346 |
| Calcium looping | £ 51 | £ 63 | £ 114 | £ 182 | £ 256 | £ 98 | £ 127 |
| Cryogenics | £ 117 | £ 165 | £ 320 | £ 481 | £ 1,011 | £ 315 | £ 363 |
| | | | | | | | |
| Median individual project abatement potential (MtCO2/yr in 2025) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | 3.0 | 0.4 | 0.8 | 0.04 | 0.2 | 0.4 | 0.1 |
| 2nd gen chemical solvent | 3.1 | 0.4 | 0.9 | 0.05 | 0.2 | 0.5 | 0.1 |
| Chilled ammonia | 3.1 | 0.4 | 0.8 | 0.04 | 0.2 | 0.4 | 0.1 |
| Potassium carbonate | 2.8 | 0.3 | 0.7 | 0.04 | 0.1 | 0.4 | 0.1 |
| Rectisol | 3.6 | 0.4 | 1.0 | 0.06 | 0.2 | 0.5 | 0.1 |
| Selexol | 3.7 | 0.4 | 1.0 | 0.06 | 0.2 | 0.5 | 0.1 |
| Calcium looping | 0.8 | 0.4 | 0.7 | 0.05 | 0.2 | 0.5 | 0.1 |
| Cryogenics | 3.3 | 0.4 | 0.8 | 0.04 | 0.1 | 0.4 | 0.1 |
| | | | | | | | |
| Median 2025 project discounted lifetime cost (£m, 10%, 15 yrs) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
| 1st gen amine solvent | £2,365 | £404 | £1,388 | £116 | £512 | £736 | £167 |
| 2nd gen chemical solvent | £1,669 | £259 | £913 | £78 | £358 | £436 | £90 |
| Chilled ammonia | £2,216 | £386 | £1,369 | £116 | £575 | £714 | £156 |
| Potassium carbonate | £4,795 | £720 | £3,214 | £268 | £1,931 | £1,678 | £287 |
| Rectisol | £2,934 | £505 | £2,574 | £239 | £1,746 | £1,351 | £244 |
| Selexol | £2,890 | £505 | £2,576 | £242 | £1,759 | £1,356 | £247 |
| Calcium looping | £332 | £197 | £708 | £71 | £374 | £375 | £77 |
| Cryogenics | £3,239 | £526 | £2,177 | £166 | £992 | £1,130 | £209 |

Other conditions as per Pragmatic Scenario

# Overall sectoral attributes for the Push Scenario

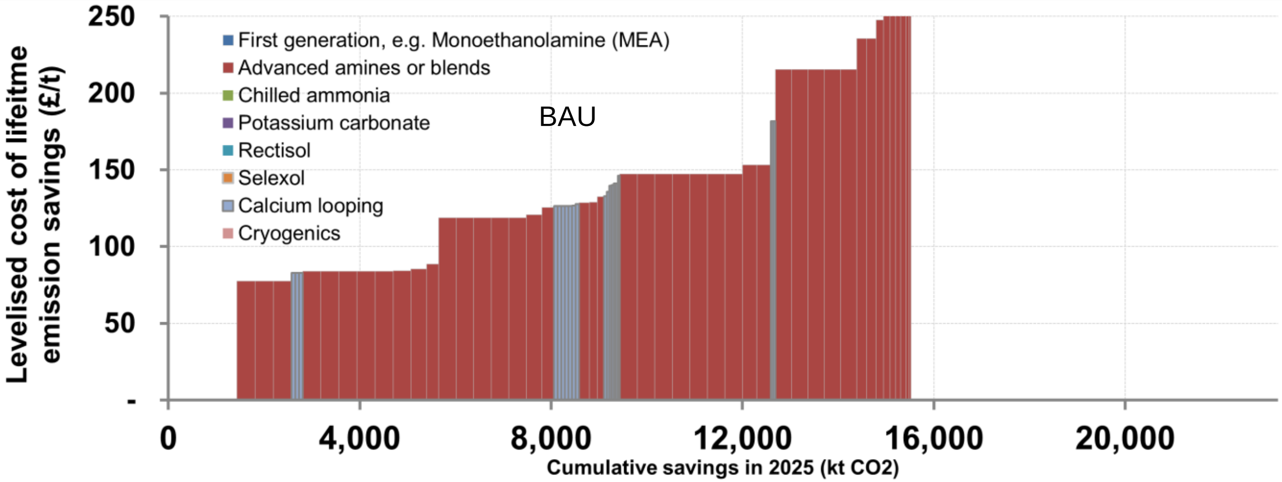
| Combined abatement potential (Mt/yr) | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 9.0 | 4.3 | 6.2 | 0.1 | 3.3 | 3.5 | 1.2 |
| 2nd gen chemical solvent | 9.3 | 4.5 | 6.5 | 0.1 | 3.5 | 3.6 | 1.3 |
| Chilled ammonia | 9.3 | 4.4 | 6.3 | 0.1 | 3.1 | 3.5 | 1.2 |
| Potassium carbonate | 8.2 | 3.9 | 5.4 | 0.1 | 2.6 | 3.0 | 1.1 |
| Rectisol | 10.8 | 5.2 | 7.7 | 0.1 | 4.3 | 4.3 | 1.5 |
| Selexol | 10.9 | 5.2 | 7.8 | 0.1 | 4.4 | 4.3 | 1.5 |
| Calcium looping | 2.3 | 4.5 | 5.3 | 0.1 | 3.4 | 3.7 | 1.3 |
| Cryogenics | 9.8 | 4.6 | 6.2 | 0.1 | 2.3 | 3.4 | 1.2 |

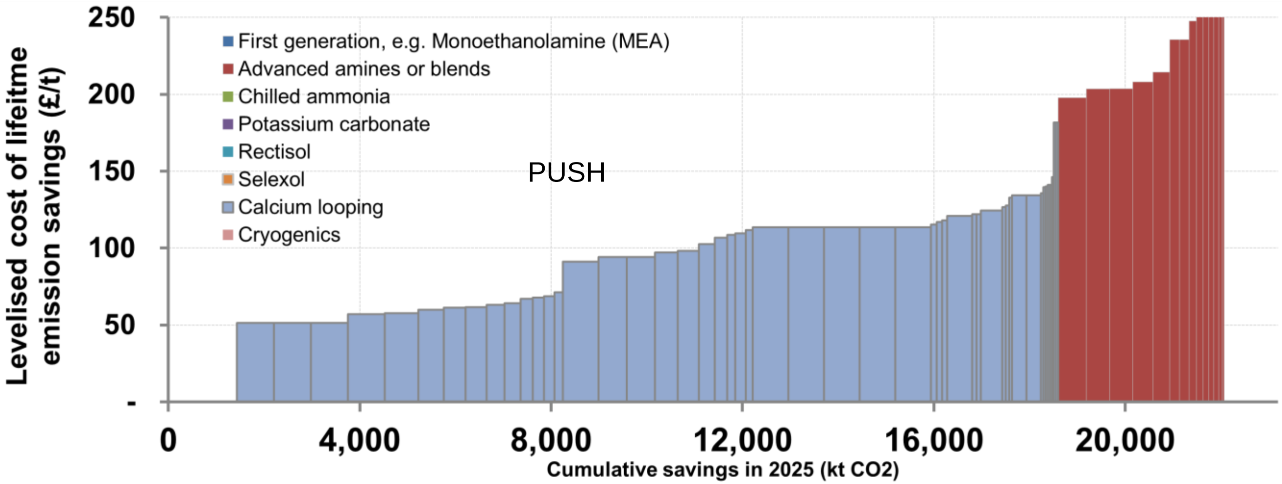
| Range of pragmatic scenario abatement costs | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 92 - 94 | 120 - 157 | 194 - 254 | 314 - 314 | 288 - 516 | 185 - 237 | 240 - 346 |
| 2nd gen chemical solvent | 63 - 64 | 77 - 93 | 123 - 153 | 202 - 202 | 198 - 327 | 108 - 129 | 129 - 171 |
| Chilled ammonia | 84 - 86 | 112 - 146 | 189 - 242 | 313 - 313 | 355 - 574 | 178 - 223 | 223 - 313 |
| Potassium carbonate | 207 - 209 | 256 - 284 | 528 - 583 | 844 - 844 | 1644 - 1940 | 515 - 559 | 528 - 615 |
| Rectisol | 96 - 98 | 130 - 156 | 297 - 340 | 513 - 513 | 877 - 1088 | 288 - 325 | 309 - 382 |
| Selexol | 94 - 95 | 128 - 155 | 293 - 338 | 514 - 514 | 868 - 1086 | 285 - 323 | 308 - 384 |
| Calcium looping | 51 - 51 | 57 - 71 | 114 - 134 | 182 - 182 | 215 - 333 | 91 - 110 | 109 - 146 |
| Cryogenics | 117 - 118 | 153 - 180 | 313 - 361 | 481 - 481 | 930 - 1168 | 301 - 340 | 325 - 402 |

| Number of projects | Steel | Cement | Refinery | Gas boiler condensing | Industrial Gas CHP | Crackers | Other chemicals |
|---|---|---|---|---|---|---|---|
| 1st gen amine solvent | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| 2nd gen chemical solvent | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Chilled ammonia | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Potassium carbonate | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Rectisol | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Selexol | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Calcium looping | 3 | 11 | 9 | 2 | 14 | 8 | 16 |
| Cryogenics | 3 | 11 | 9 | 2 | 14 | 8 | 16 |

Other conditions as per Pragmatic Scenario

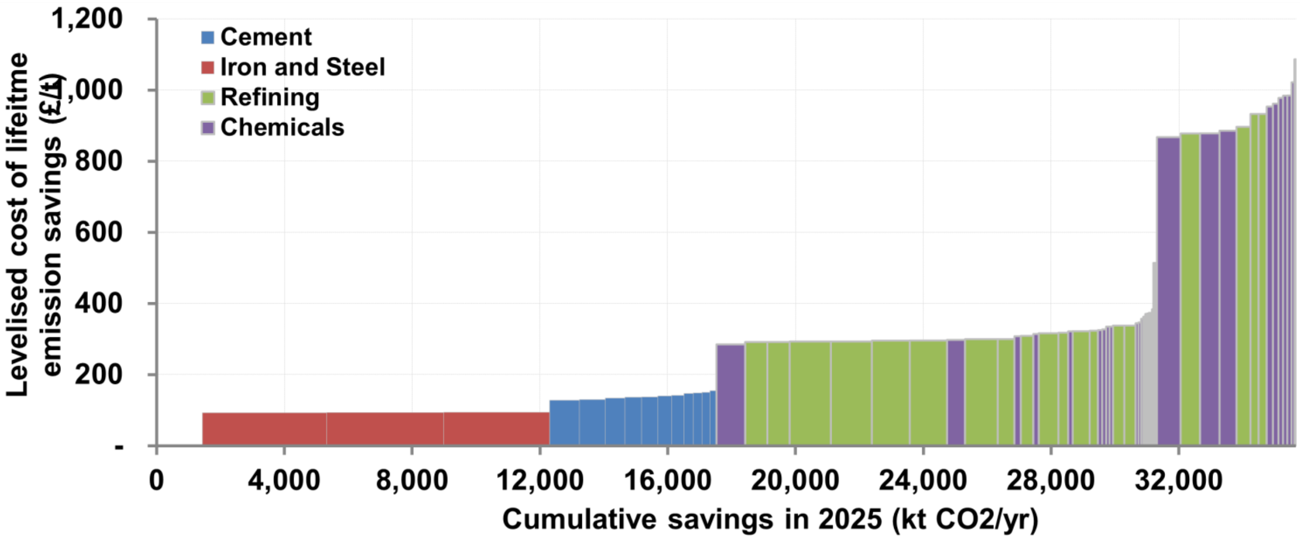
# Least cost technologies (in £/t abated) in BAU and Push scenarios

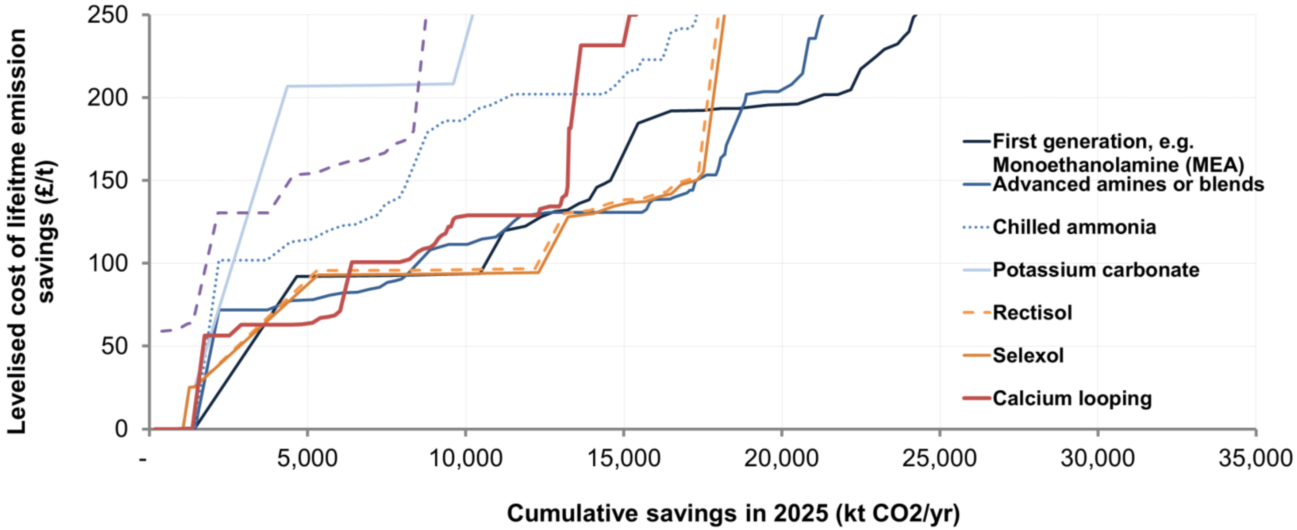



Other conditions as per Pragmatic Scenario

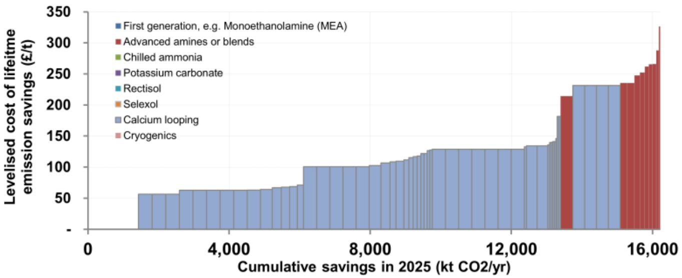
# Impact of choosing projects with maximum $CO_2$ abatement potential

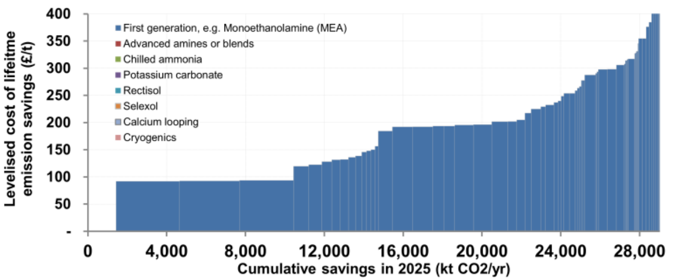


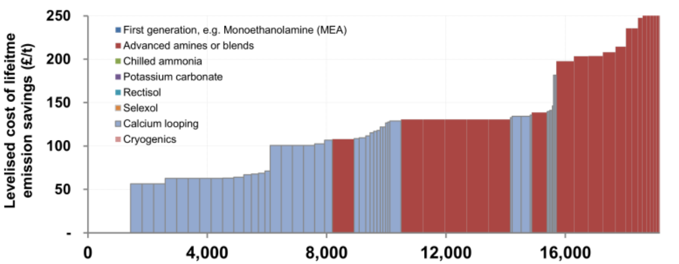



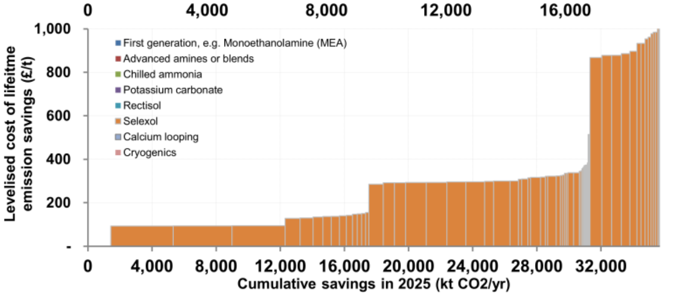

Other conditions as per Pragmatic Scenario, where projects are chosen on the basis of minimal £/$tCO_2$ abated

# The model shows sensitivity of technology choice and total $CO_2$ abated to assumptions on investor priorities.




| Investor priority | Favoured technology |
|---|---|
| Minimise project discounted lifetime cost (in £) | Calcium looping projects dominate 2025 MACC |
| Maximise $tCO_2$/yr captured OR choose highest technologies with highest TRL | 1st generation amines (e.g. MEA) projects dominate 2025 MACC |
| Minimise £/$tCO_2$ abated for each project | Calcium looping and 2nd generation amines dominate 2025 MACC |
| Maximise $CO_2$ abatement potential | Physical solvents (e.g. selexol) dominate MACC |

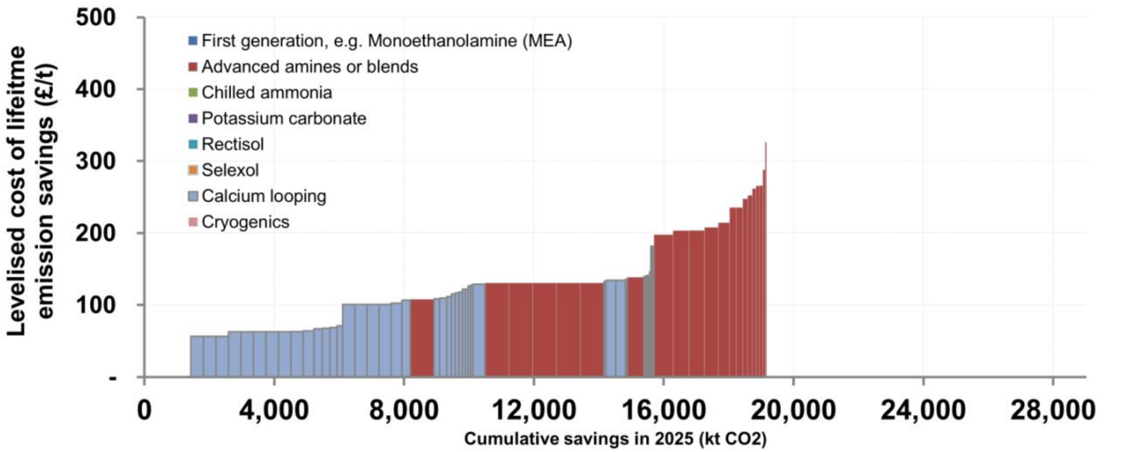
**$MtCO_2$/yr abated**

Other conditions as per Pragmatic scenario

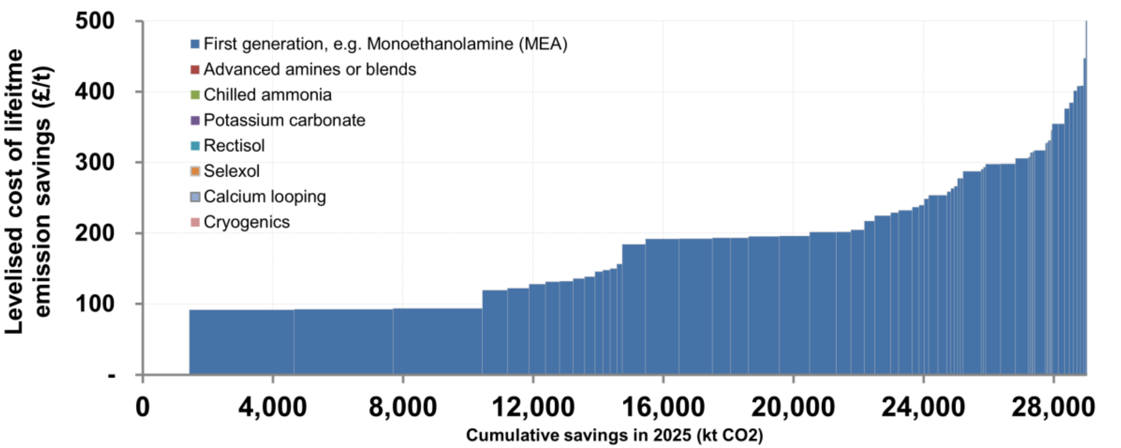
# Sensitivity comparison of 2025 abatement potential for different assumptions on technology availability

- Compared to high TRL technologies (e.g. MEA), the currently lower TRL technologies can be deployed at lower levelised costs, but the potential scale of individual projects is smaller
- Excluding these technologies as an option for 2025 increases individual project costs, but high TRL technologies have the potential to be deployed at larges scales.

All capture technologies available for 2025 scenario

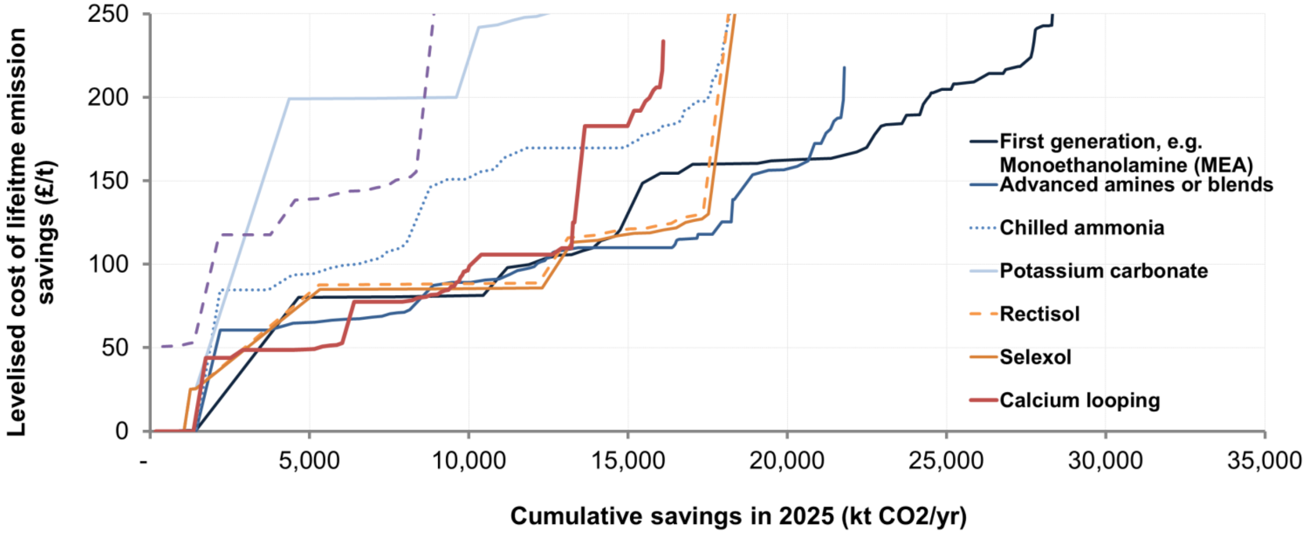



Only currently mature technologies (first generation amines, selexol and rectisol), available in 2025

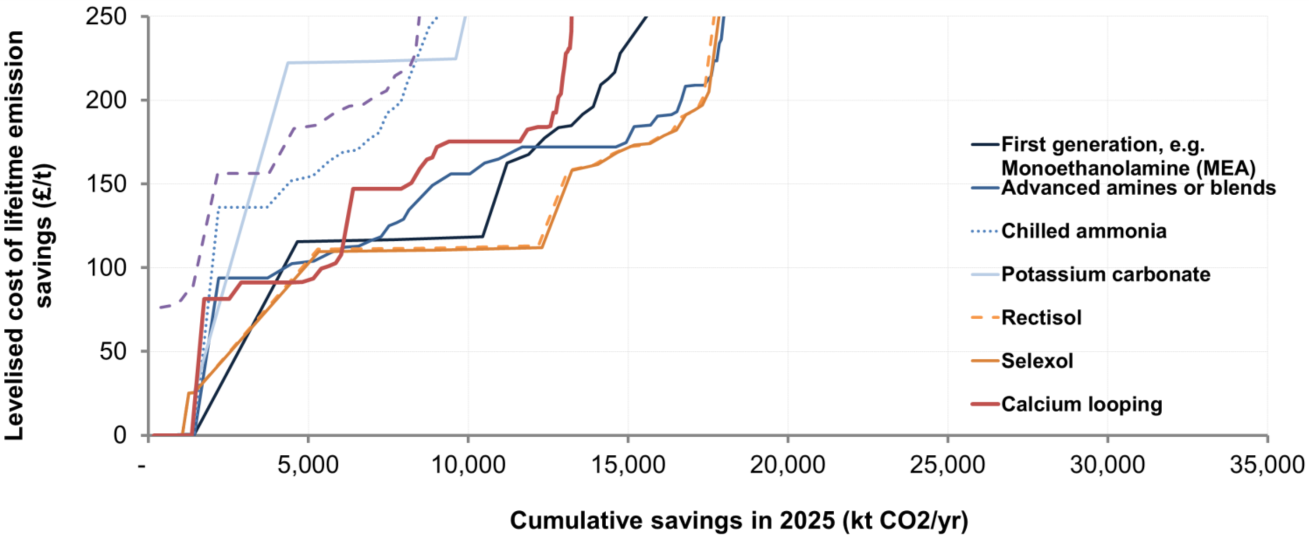



*Pragmatic scenario (e.g. technology selection ased on lowest project levelised costs)

# Capture capex sensitivity – technology MACC comparison

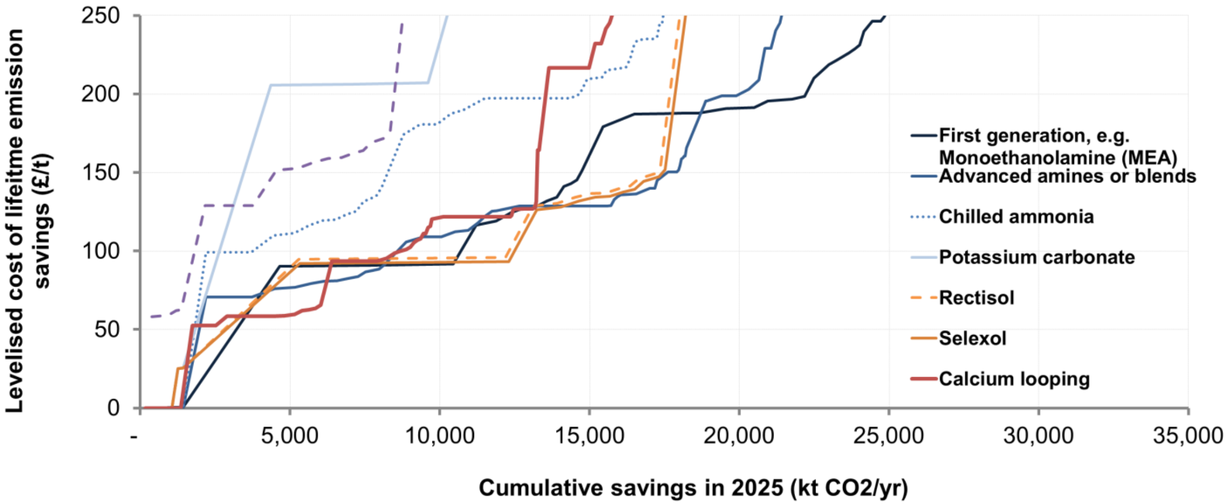



Reference Capture capex = 50% of Pragmatic scenario

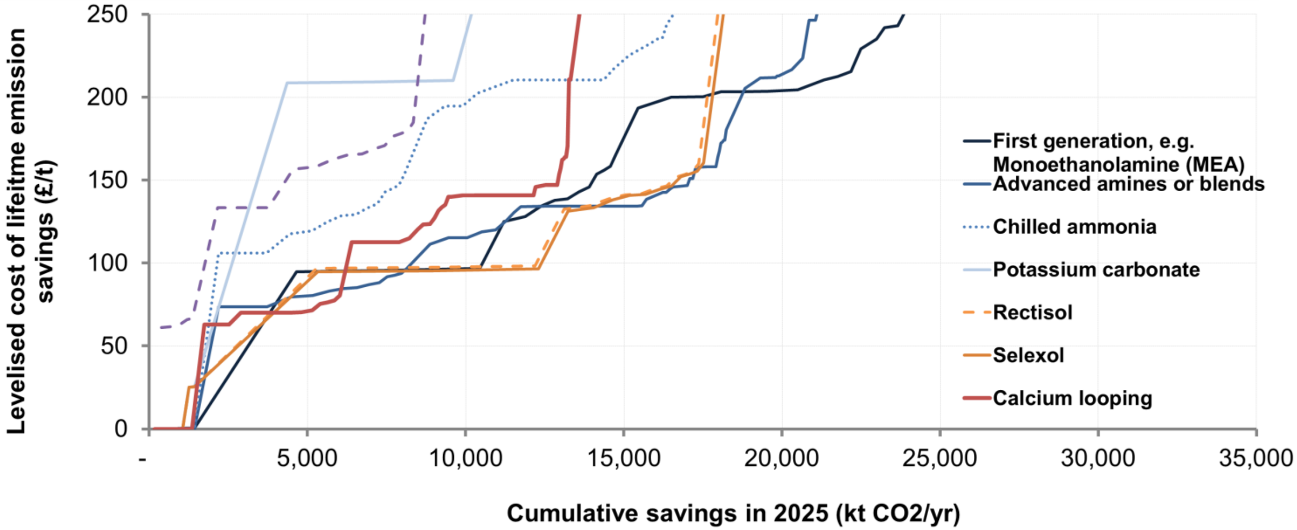



Reference Capture capex = 200% of Pragmatic scenario

# Capture fixed opex sensitivity – technology MACC comparison

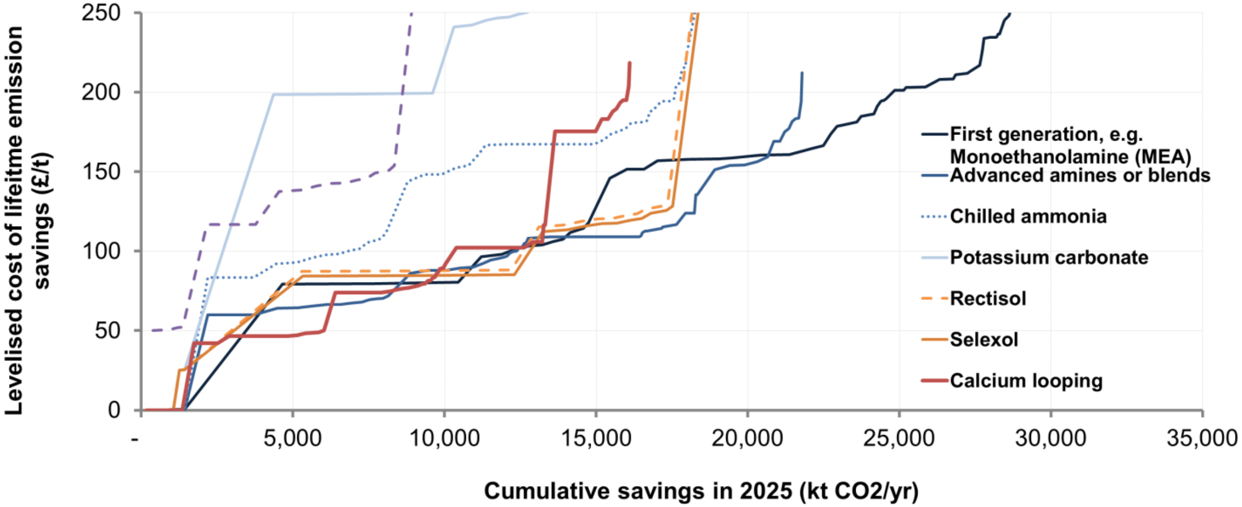



Low fixed opex scenario

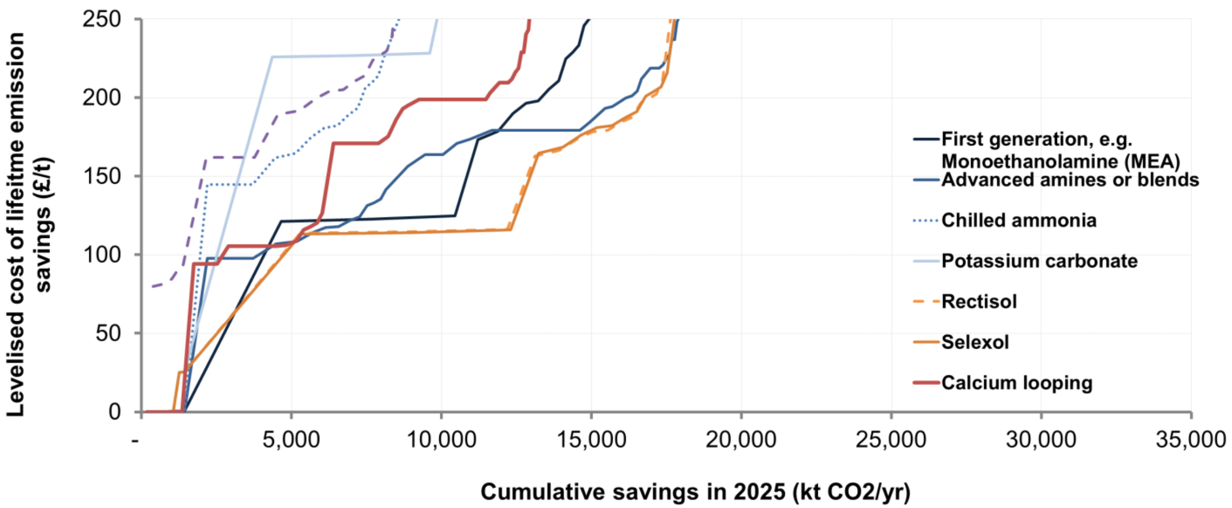



High fixed opex scenario

# Combined impact of capture capex and opex uncertainty – technology MACC comparison

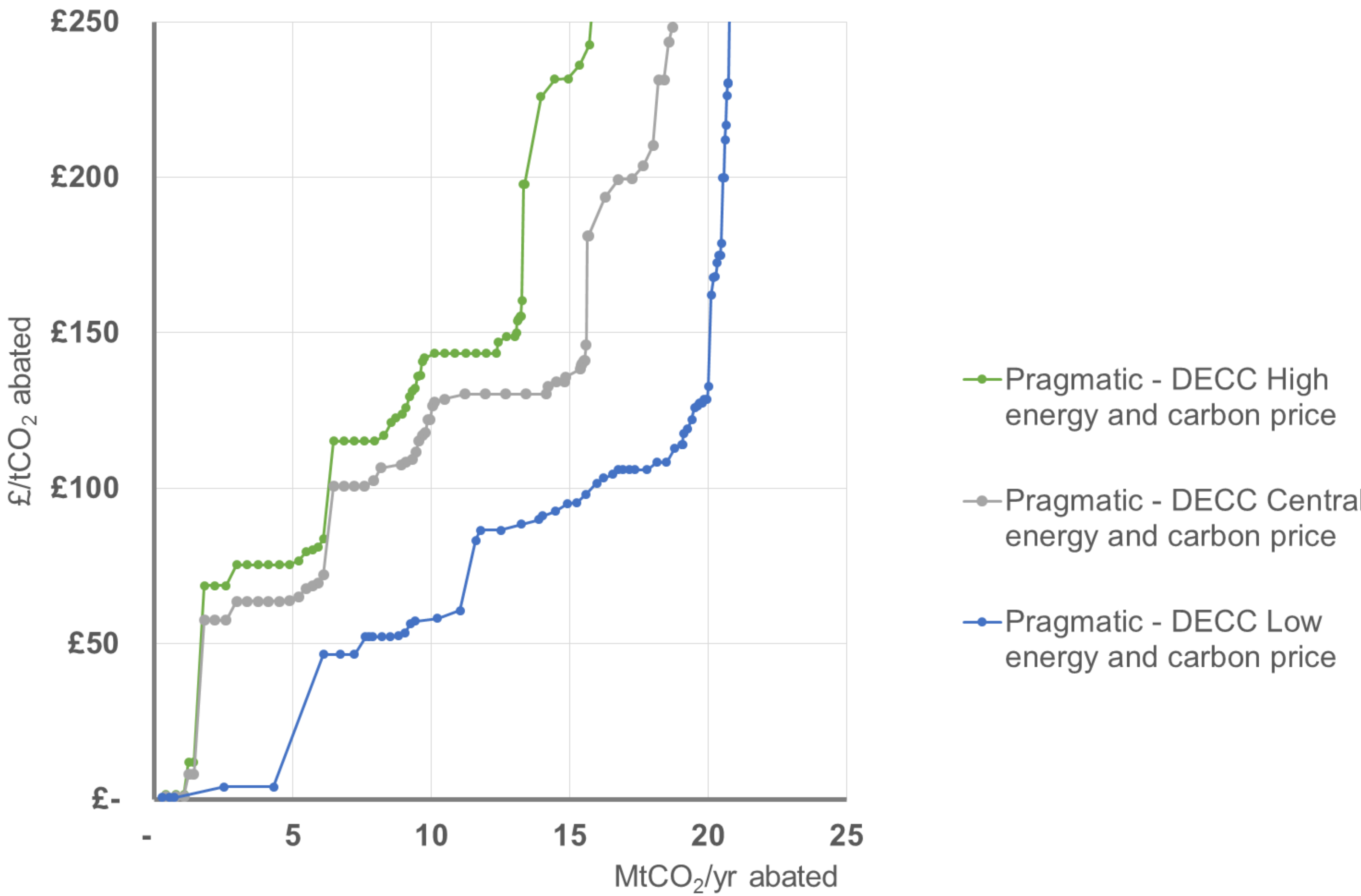



Low cost scenario (-50% cf. pragmatic)




High cost scenario (+100% cf. pragmatic)

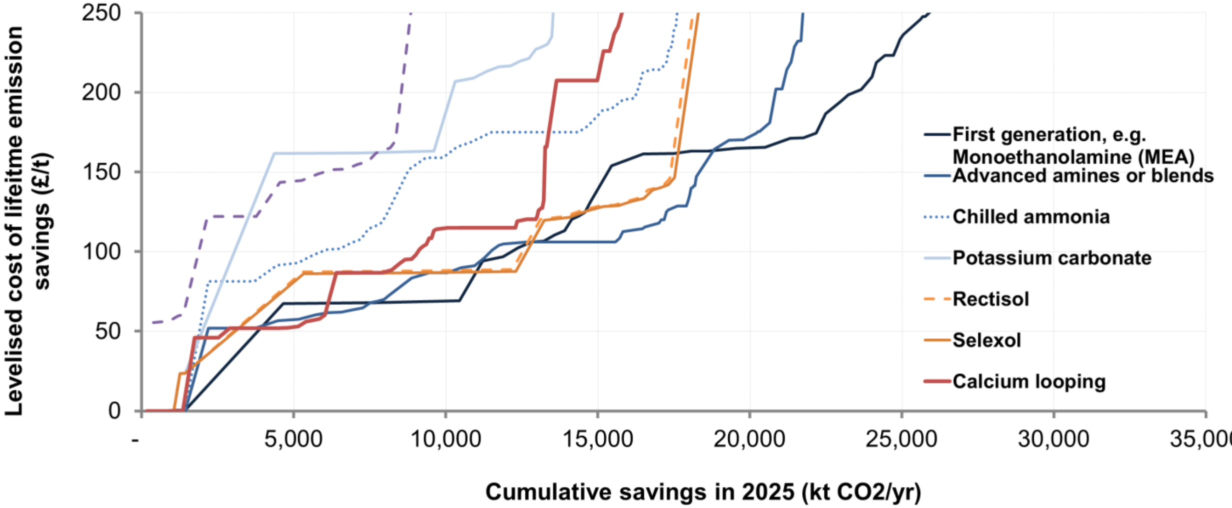
# Impact of energy and carbon prices – overall MACC based on projects with lowest £/t$CO_2$ abated



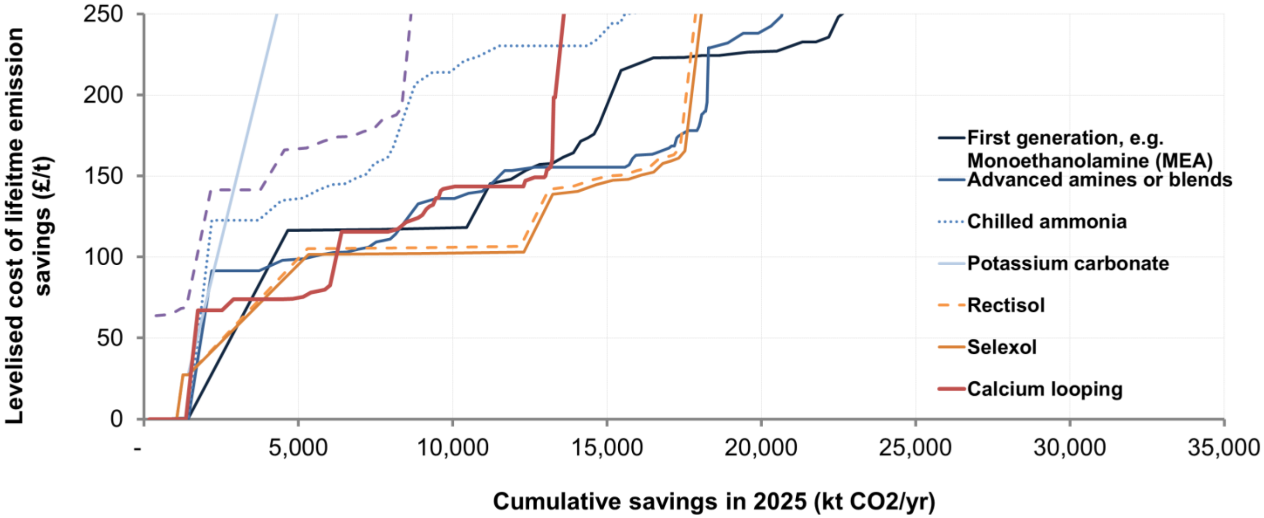

The techno-economic model purely shows costs. Note that no assumption made on avoided ETS or carbon price *revenues* within the abatement cost calculation Here high carbon price implies high costs associated with $CO_2$ payments for an on-site boiler to supply steam for capture plant.

Other conditions as per pragmatic scenario. Costs exclude post-capture $CO_2$ compression, transport or storage.

# Impact of energy and carbon prices – technology comparison



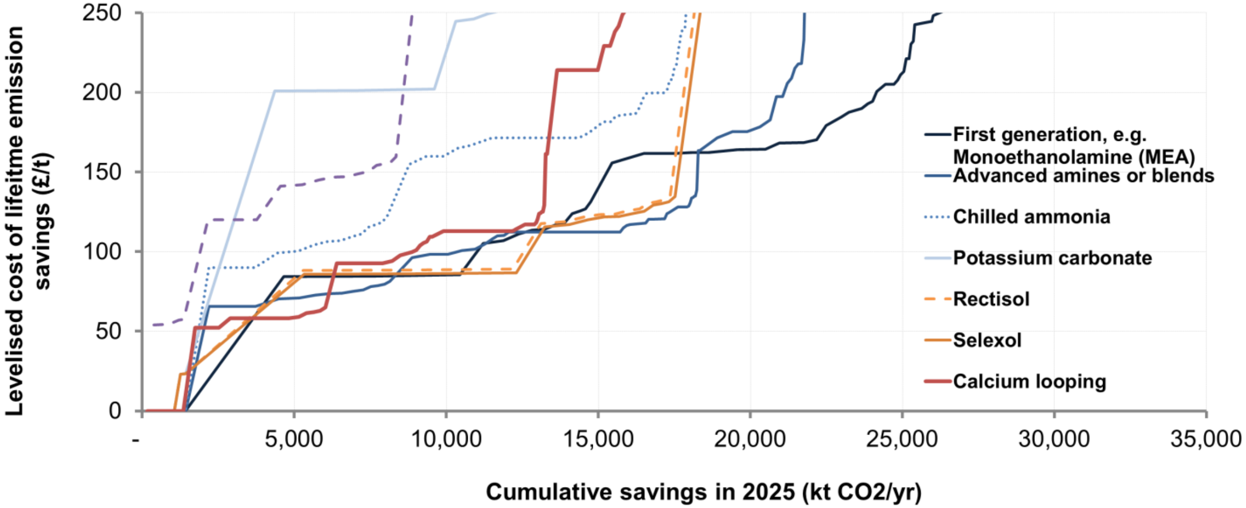

Low prices scenario



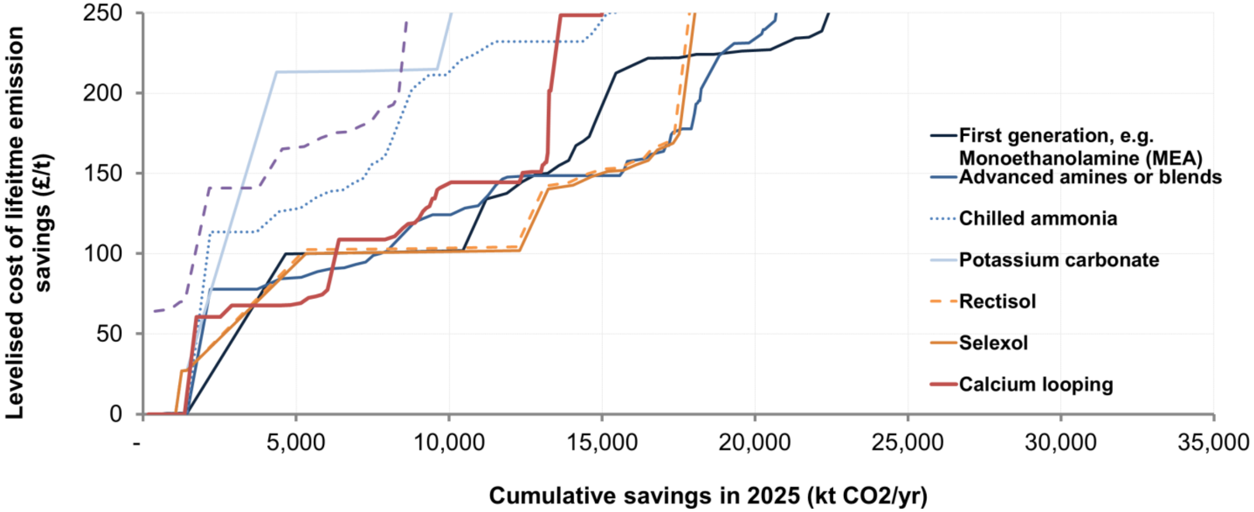

High prices scenario

Other conditions as per pragmatic scenario

# Impact of discount rate – technology comparison



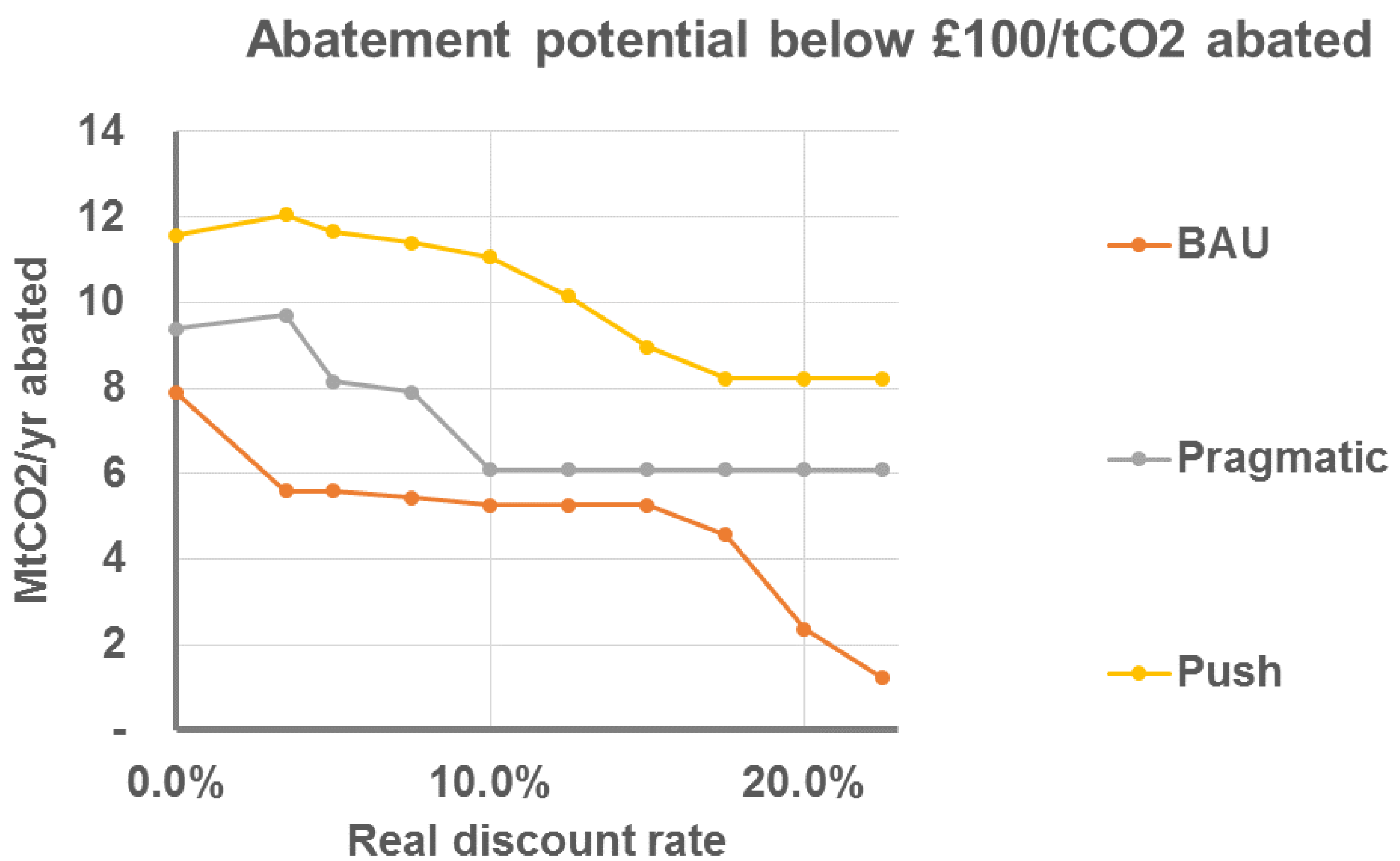

3.5% Discount rate




15% Discount rate

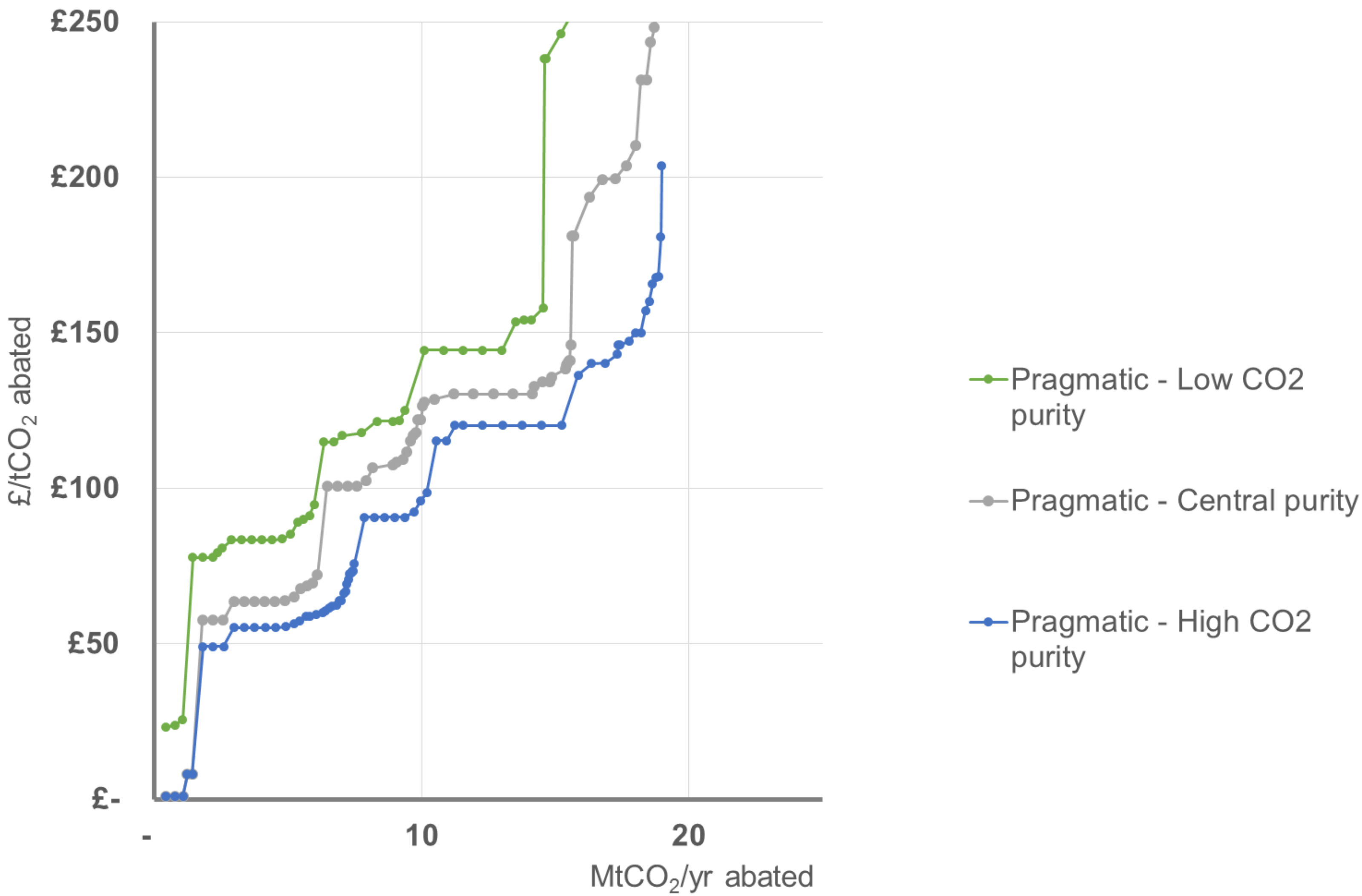
Other conditions as per pragmatic scenario

# Impact of discount rate – impact of technology deployment rate



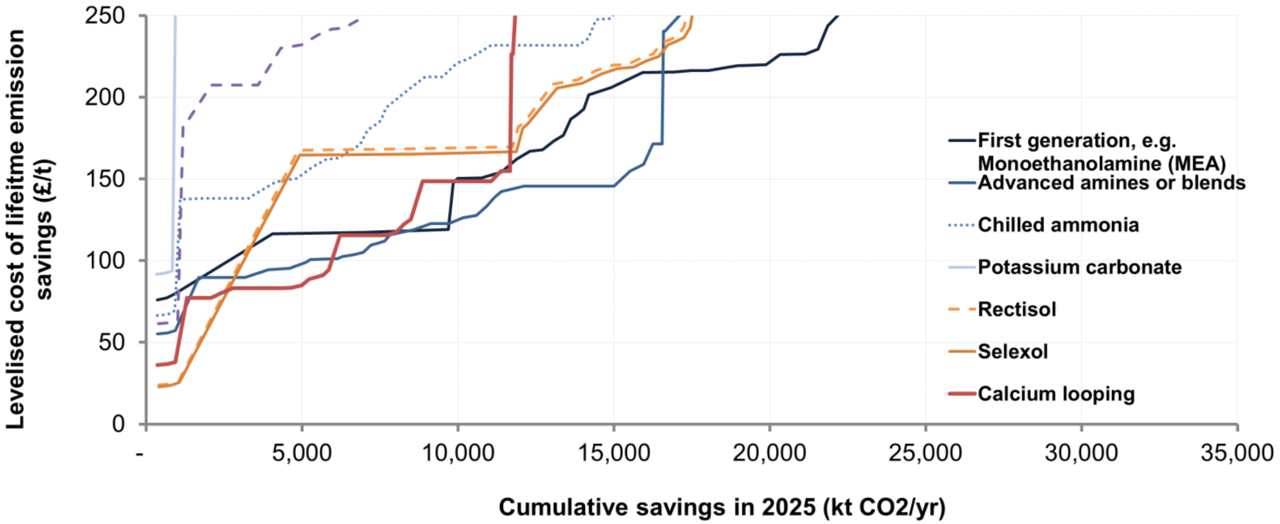

Other conditions as per pragmatic scenario

# Impact of source $CO_2$ purity

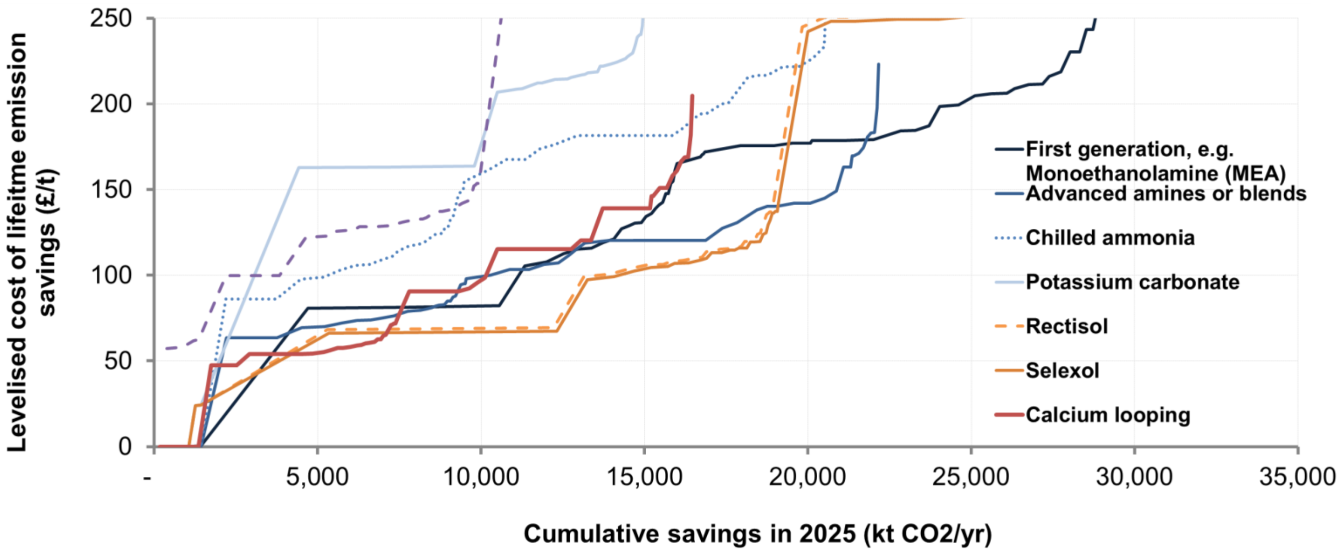



Other conditions as per pragmatic scenario

# Source $CO_2$ purity most strongly impacts the cost effectiveness of cryogenics, physical solvents and potassium carbonate



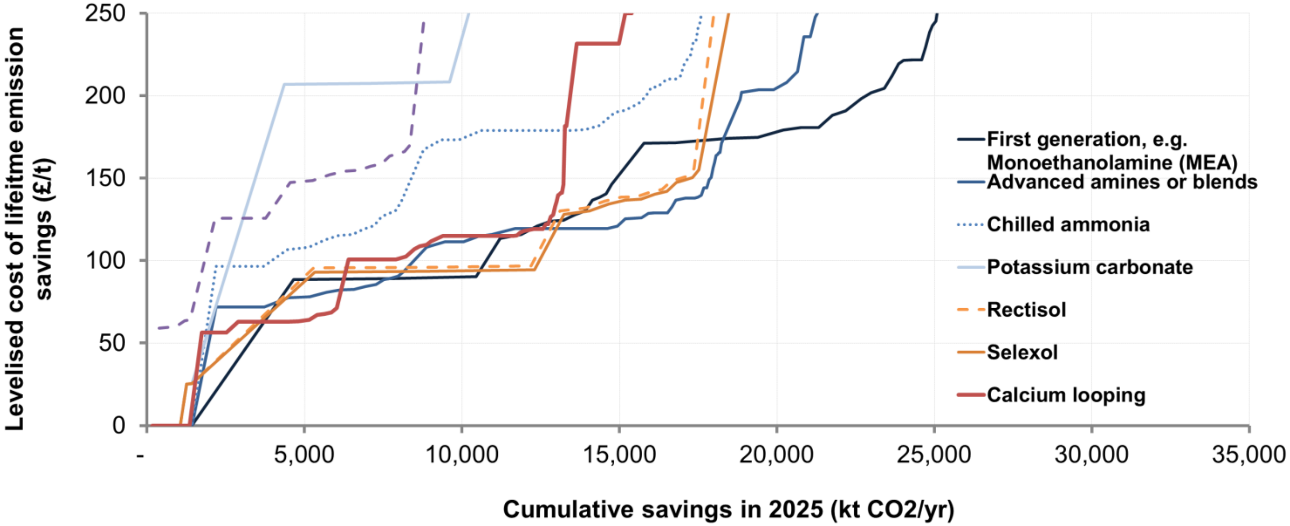

Low source $CO_2$ purity




High source $CO_2$ purity

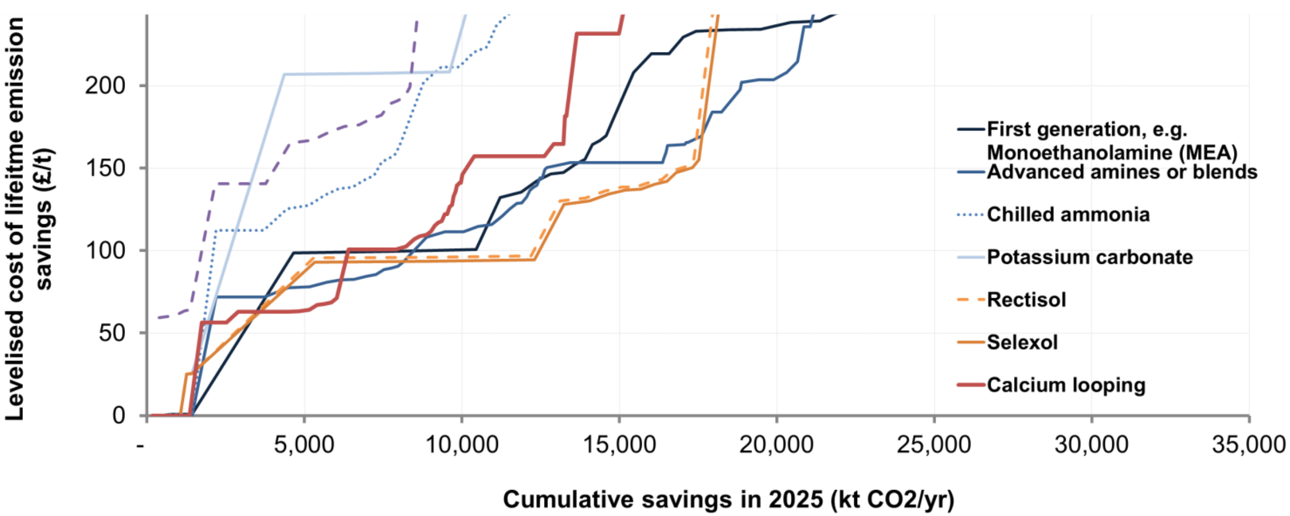
Other conditions as per pragmatic scenario

# Sensitivity of pragmatic scenario to pre-treatment capital and operating cost assumptions




<u>Low pre-treatment cost</u>

Capital and operating costs of pre-treatment are half of those in the Pragmatic Scenario

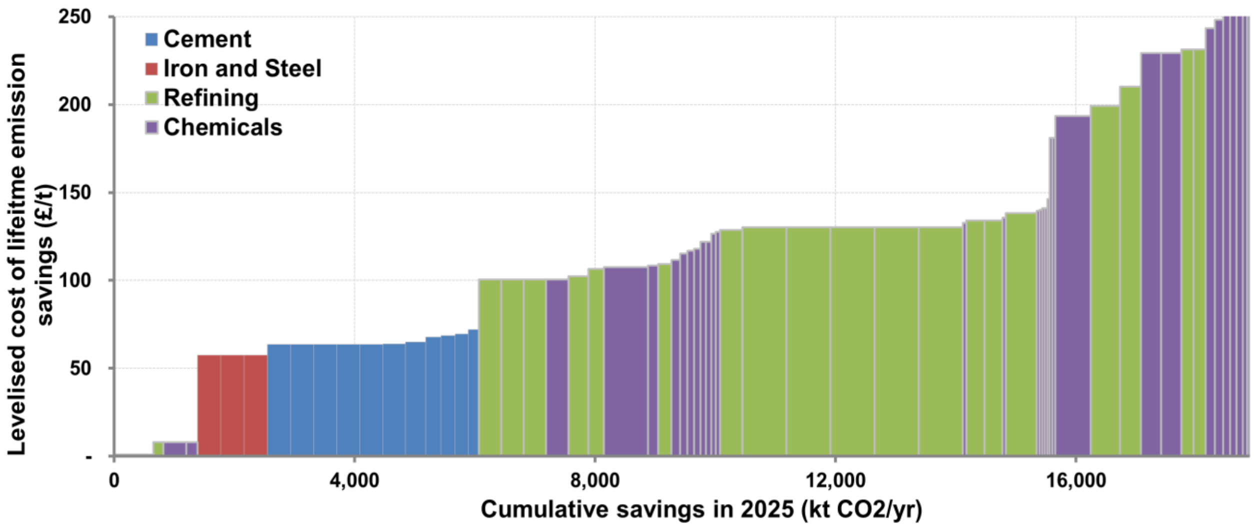


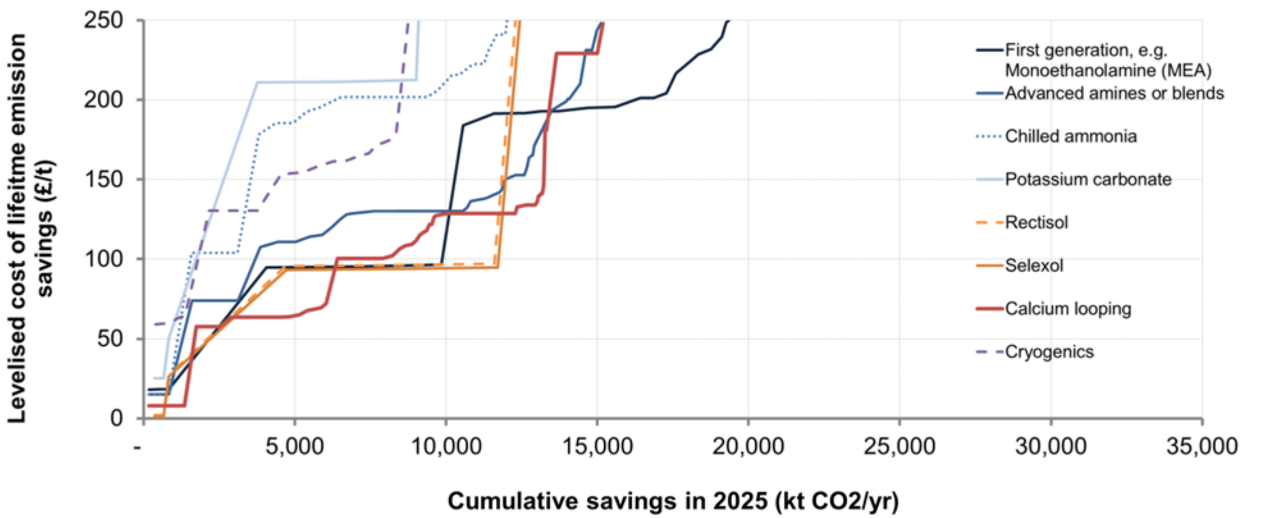

<u>High pre-treatment cost</u>

Capital and operating costs of pre-treatment are 2x those in the Pragmatic Scenario

Other conditions as per pragmatic scenario

# Moderate restrictions based on site current COMAH status should have limited impact.

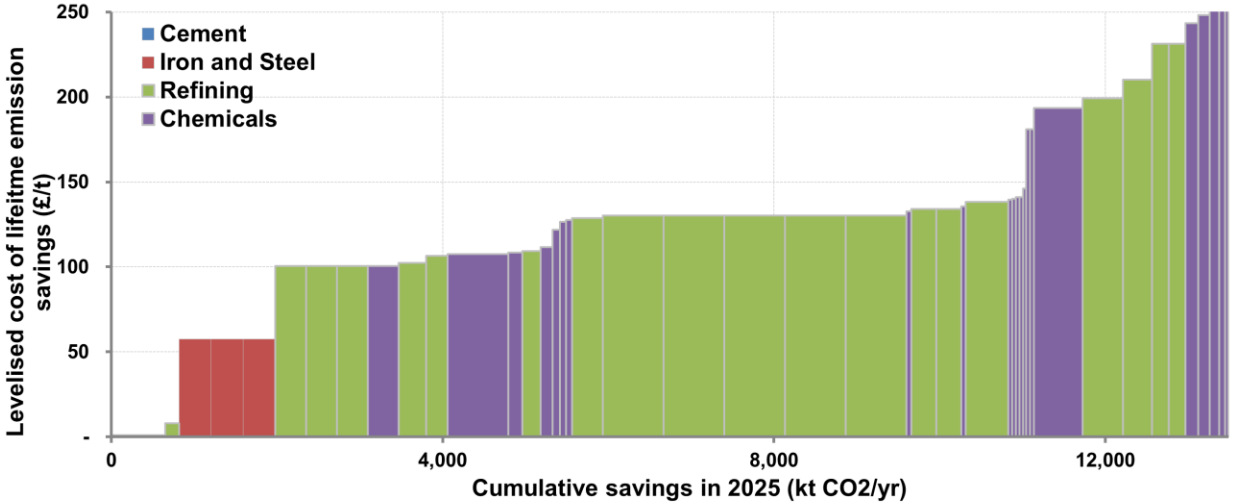


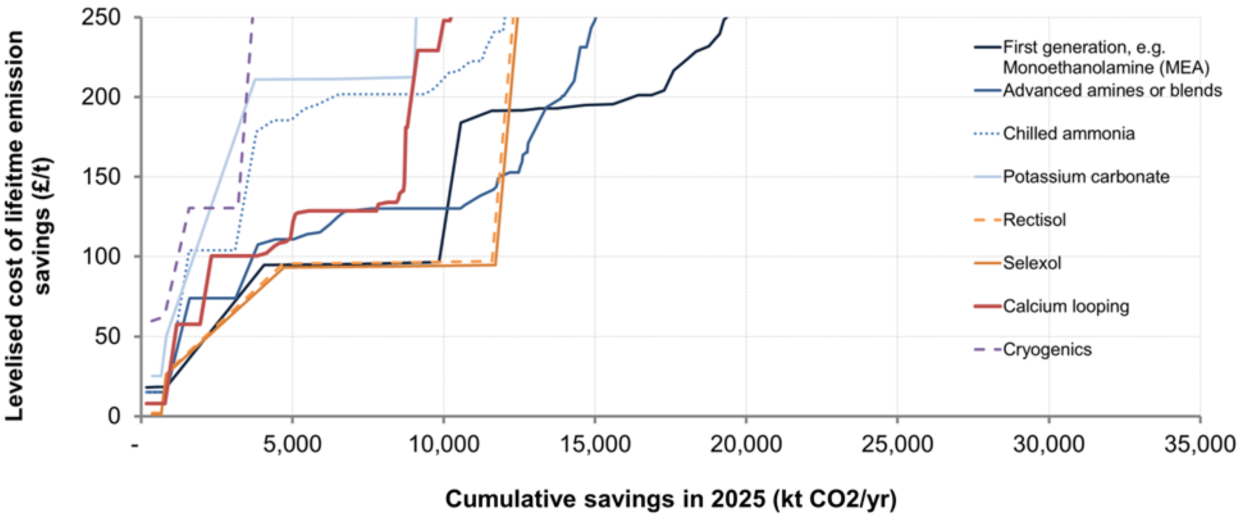




| Technology | Scenario for Assumption Minimum COMAH |
|---|---|
| 1st gen amine | Top |
| 2nd gen amine | Lower |
| Ammonia | Top |
| Potassium carbonate | Lower |
| Rectisol | Lower |
| Selexol | Lower |
| Calcium looping | None |
| Cryogenics | None |

Other conditions as per pragmatic scenario, where no COMAH restriction is assumed

# Impact of limiting $CO_2$ capture to sites with Top Tier COMAH Status only already could significantly restrict cement CCS in 2025.

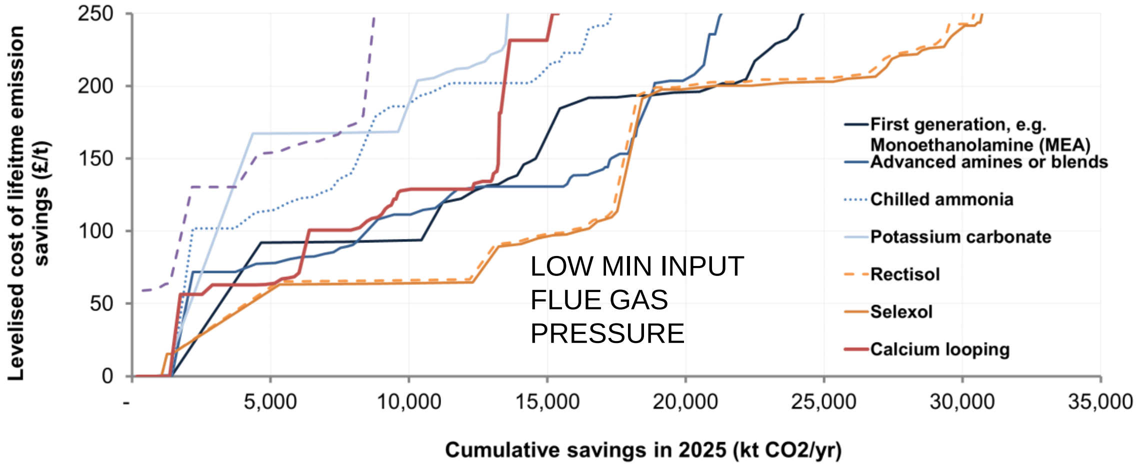


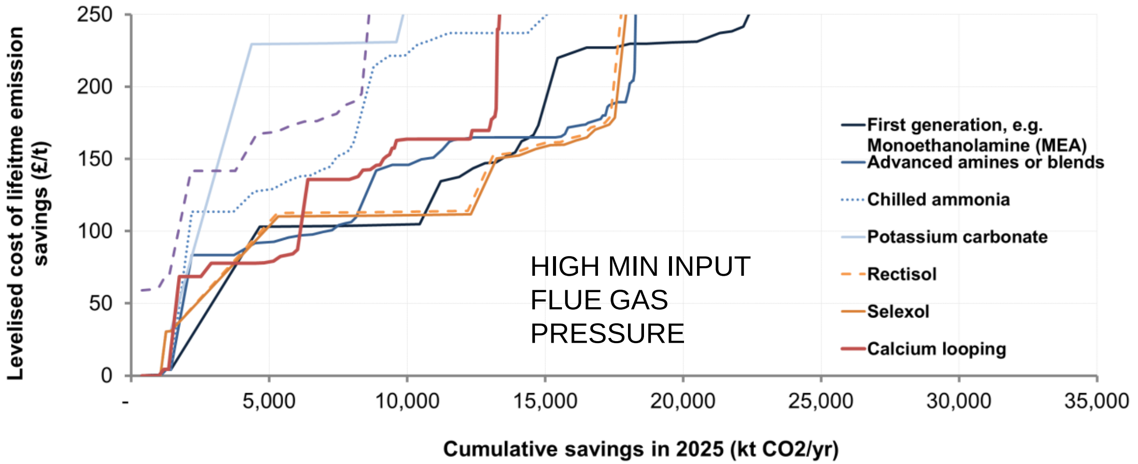




| Technology | Scenario for Assumption Minimum COMAH |
|---|---|
| 1st gen amine | Top |
| 2nd gen amine | Top |
| Ammonia | Top |
| Potassium carbonate | Top |
| Rectisol | Top |
| Selexol | Top |
| Calcium looping | Top |
| Cryogenics | Top |

Other conditions as per pragmatic scenario, where no COMAH restriction is assumed.

# The cost effectiveness of potassium carbonate and physical solvent based capture is highly sensitive to initial flue gas pressure.




| | Minimum flue gas pressure for operation | | |
|---|---|---|---|
| Technology | Low | Pragmatic | High |
| 1st gen amine | 1 bar | 1 bar | 2 bar |
| 2nd gen amine | 1 bar | 1 bar | 2 bar |
| Ammonia | 1 bar | 1 bar | 2 bar |
| Potassium carbonate | 10 bar | 30 bar | 50 bar |
| Rectisol | 10 bar | 30 bar | 50 bar |
| Selexol | 10 bar | 30 bar | 50 bar |
| Calcium looping | 1 bar | 1 bar | 2 bar |
| Cryogenics | 1 bar | 1 bar | 2 bar |

Other conditions as per pragmatic scenario

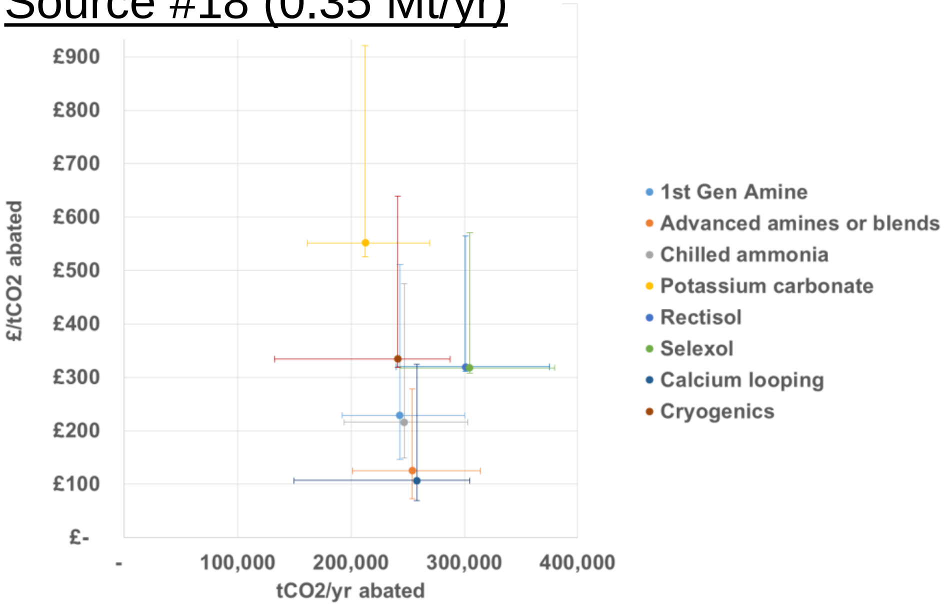
# Exploring the uncertainty in cost and capacity for technologies: Example from Grangemouth refinery

| Source Parameter | Source#18 "Grangemouth Cracker" | Source#20 "Grangemouth refinery excl. cracker or CHP" |
|---|---|---|
| Input stream | 0.35 $MtCO_2$/yr | 1.6 $MtCO_2$/yr |
| Source archetype | Cracker | Refinery |
| Input $CO_2$ concentration | 10% | 10% |
| Flue gas pressure | 1 bar | 1 bar |
| Vent complexity | Many vents | Many vents |
| NOx | 100 ppm | 600 ppm |
| SOx | 100 ppm | 1200 ppm |
| Technical capacity | High | High |
| Water availability | High | High |
| COMAH status | Top | Top |
| Waste heat available | No | No |

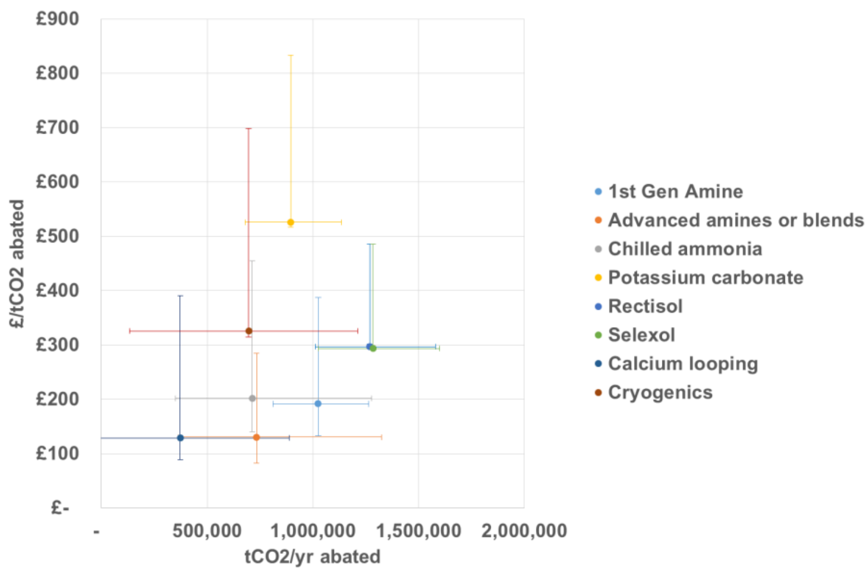
| | Pragmatic scenario | "Best" case | "Worst" case |
|---|---|---|---|
| Energy/ carbon prices | DECC Central | DECC Low | DECC High |
| Capture capex and opex multiplier | 1x | 2x baseline | 0.5x baseline |
| $CO_2$ source purity | 10% | 12% | 8% |
| Source $MtCO_2$/yr production in 2025 | Baseline | 125% of baseline | 80% of baseline |
| Technology development | Pragmatic | High | Business as Usual |
| Lifetime | 15 yrs | 20 yrs | 10 yrs |
| Real discount rate | 10% | 10% | 10% |

# The model allows the cost and capacity uncertainties for all technologies to be compared systematically for different sources.

Source #18 (0.35 Mt/yr)



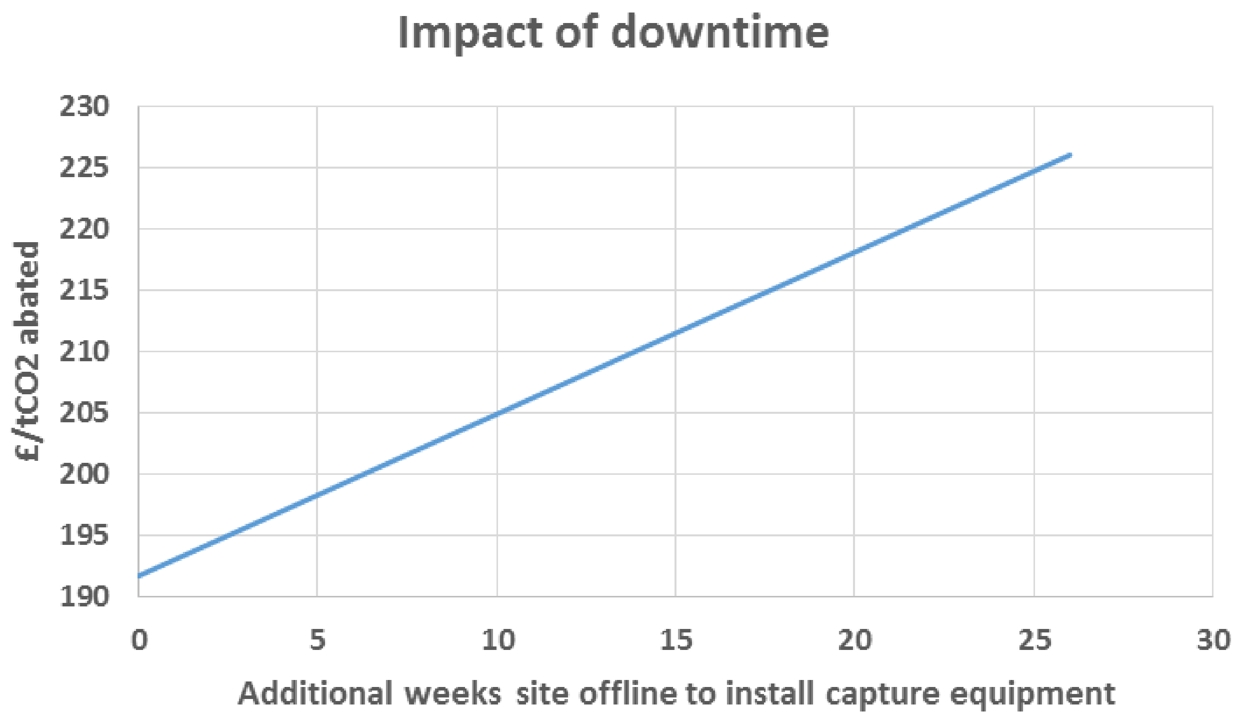

Source #20 (1.6 Mt/yr)




- Of technologies available, 1st generation amine solvents (e.g. MEA) should have low £/t$CO_2$ abated costs and high abatement potential.
- If developed at scale, either calcium looping or 2nd generation amines should have lowest £/t$CO_2$ abated costs, however, the capacity (i.e. Mt/yr) that these can be implemented in the period 2020 to 2025 is uncertain and will likely be significantly smaller than for 1st generation amines.
- Conventional potassium carbonate, physical solvents or cryogenic technologies require significant electricity consumption and are predicted to be more expensive to implement, unless as part of novel configurations.

# The importance of installing capture equipment around the time of major site refurbishment

- Example Grangemouth refinery source #20 with 1st gen amine capture technology.
- Base case in pragmatic scenario has £192/t$CO_2$ abated, excluding downtime costs.
- What if each day additional downtime results in an effective loss of £1million*?
- Impact of additional downtime is to increase the effective cost of capture as shown below (modelled as an effective increase in capex of £7m/week).




* A. Roberts, UKPIA, Personal Communication

# Outline

- Overall Project Methodology
- $CO_2$ capture technologies
- $CO_2$ sources
- Techno-economic analysis of industrial $CO_2$ capture
- Process simulation case studies
- $CO_2$ utilisation review

# Process simulation complements the techno-economic modelling

- Whereas techno-economic model provides a “top down” perspective on overall project costs and benefits, to help identify the most relevant capture technologies, the process simulation provides a “bottom up” perspective on how capture might be implemented at actual UK industrial plants in the period to 2025.
- The literature supporting techno-economic studies is frequently opaque and it is difficult for stakeholders to have a clear understanding of cost boundaries. In contrast process simulation provides a transparent and detailed description of the key capture infrastructure (including sizing of key components) and accompanying mass and energy flows.
- Process simulation helps understand the breakdown and sensitivities of capital and operating costs, and thereby understand priorities for technology development, and management of costs and risks.
- The underlying cost databases for process simulation tools generally require numerous assumptions to convert “equipment costs” into total installed costs.
- Process simulation models are very resource intensive to develop and analyse, therefore very few scenarios can be examined.
- The choices of source and technology to model were based on a combination of results from the techno-economic modelling (i.e. projects with low cost or high abatement potential in 2025), consideration of sites with plausible opportunities for accessing $CO_2$ transport and storage infrastructure in 2025, and discussions with sector trade associations and individual companies.

# Energy price assumptions in process simulation

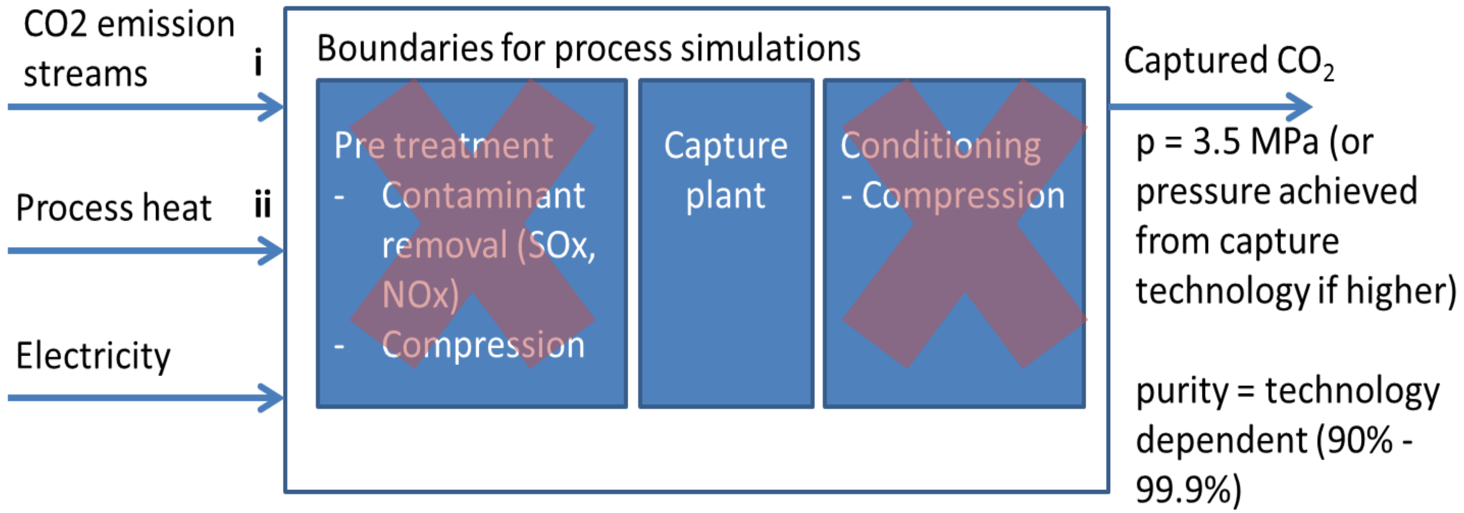
- The energy cost assumptions for the process simulation were prepared using effective annual prices for gas and electricity as follows, assuming a discount rate of 10%.

| Effective fuel price (p/kWh assuming discount rate of 10% and lifetime of 15 yrs) | Pilot (assumed 2020) | Demo scale 2025 |
|---|---|---|
| **Electricity** | 7.32 | 7.57 |
| **Gas** | 1.77 | 1.78 |

# Four case studies on process simulation

1. Grangemouth oil refinery with MEA capture
2. Lafarge Tarmac Dunbar cement works with MEA capture
3. Tata Scunthorpe iron and steel plant with MEA capture at CHP unit
4. GrowHow ammonia production site with $CO_2$ compression

# Scope of study




**i** - only process emissions (excluding power and heat production emissions)
- in case multiple stack emissions are combined, piping etc to bring together is included

**ii** - including integration cost in case of process heat recovery
- assume natural gas boiler if no process heat is available (default assumption)

Further Notes
- Compression plant not simulated
- Pre-treatment processes simulated but not costed

# Detailed assumptions

- Challenges in conveying and blending streams from various flue sources are not considered within the process simulation, which assumes that the flue gases can be combined to give a gas with weighted average $CO_2$ concentration. The costs of initial flue gas collection pipelines are not modelled in the process simulations.
- A basic MEA $CO_2$ capture plant configuration is selected for the study (complex configurations such as split-flow considerations are ignored)
- An optimal lean loading[1] of 0.28 mol $CO_2$/mol MEA was assumed for the scenarios. This may introduce some inefficiencies in the performance. A full optimization for each scenario would require substantial development work.
- A rich loading of about 0.473 was assumed to be representative of the physical constraint on the capacity of the MEA solvent in chemical absorption processes.
- The column models were rate-based, distributed models. A trade-off between simulation performance (times) and accuracy was made reducing the number of discretization elements. This resulted in slight mass imbalances for certain components of up to about 1%.
- $O_2/N_2$ lumped as inert material
- $SO_X$ and $NO_X$ compositions are not considered at capture plant boundary
- Solvent degradation effects ignored
- As degradation rates are not estimated, no reclaimer is modelled or costed in this unit.

[1] Lean Amine Loading (LAL) is determined by measuring the amount of acid gas contained in the amine stream exiting the Amine Regenerator, expressed as a mol ratio of $CO_2$ and amine; (mol of CO2 + mol H2S)/mol amine.

# Detailed assumptions

- For some sensitivity cases (those that use the simplified capture plant), certain process/design variables were assumed to be the same as the baseline scenario. Only the following were updated by the simplified capture model:
  - Absorber column diameter (height is assumed to be same as the baseline)
  - Stripper column diameter
  - Steam flowrate to the reboiler
  - Reboiler heat duty
- Fuel and electricity costs are based on DECC's 2025 costs for demos.
- All other costs are based on Q3 2013. These could be translated to the same basis by assuming an appropriate discount rate.
- Where more than one train of capture plant is required, all costs provided are overall costs except those provided in the equipment list.

# Detailed assumptions

The estimate of total fixed capital costs is based on the hand factor methodology.

- This is an established cost estimate methodology in the process industry. The methodology provides factors to estimate the total capital costs based on the equipment purchase costs.
- The total capital cost consists of four main components;
  - supply of equipment
  - supply of materials
  - transport and installation
  - indirect costs.

  Each of these is again broken down in subcomponents.
- The hand factors depend on the type of equipment. For this analysis an overall hand factor of 3.88 is used (eg total capital cost is 3.88 times the equipment purchase cost), based on those for columns, as these make up the bulk of the equipment costs.
- Engineering and design costs of 30% are included, and a further typical contingency is included of 30%.

# Inputs

- 30 wt% MEA solvent
- 90% capture target
- >95% purity $CO_2$ by volume
- Absorber operating at atmospheric pressure
- Mellapak 250Y structured packing used in Absorber and stripper columns
- Stripper feed temperature ~ 102°C
- Stripper operating pressure ~ 1.67bara
- Steam pressure 3.5bara
- Cooling water temperature is assumed to be 10°C
- 8400 hours of continuous operation assumed in a year

# Key elements of process design (1/2)

1. All emissions from relevant sections are blended and treated in a common pre-treatment area. The pretreatment area consists of a number of unit operations.
2. A Selective Catalytic Reduction unit is used to capture 90% of NOx emissions.
3. An Electrostatic precipitator is used to reduce virtually all the particulate emissions.
4. A blower raises the pressure of the gas stream to overcome the pressure drops of the downstream systems (FGD, DCC  and absorber)
5. A Flue gas desulphurisation (FGD) unit is required to reduce SOx emissions to acceptable levels. Some carbon dioxide is generated in the process and affects the mass balance. This is seen in the difference in the mass flows of $CO_2$ in Streams 1 and 2.

# Key elements of process design (2/2)

7. The flue gas is cooled in a direct contact cooler (DCC) to about 40°C. The gas is saturated with water vapour at those conditions.
8. The gas flows into the absorber where it is counter currently contacted with MEA solvent. MEA chemically absorbs $CO_2$ and the resultant (rich) solvent is pumped from the absorber sumps through a lean/rich heat exchanger to the stripper column for regeneration. The duty required in the reboiler of the column for solvent regeneration is supplied by low-pressure steam (about 3.5bara). In the partial condenser of the stripper column, $CO_2$ is separated from water vapour before it is compressed in the downstream compressor units.
9. The hot regenerated (lean) solvent heats up the cooler rich solvent in the lean/rich heat exchanger and is further cooled in a lean amine cooler before it flows to the absorber, completing the cycle.
10. A buffer tank is used in the process where make-up solvent and/or water could be added to the process. The tank also provides additional flexibility to the process based on its capacity.
11. Gas from the top of the absorber is cooled and vapourised amine solvent is captured in the absorber wash water section.

# Scenarios for process simulation for the refinery sector

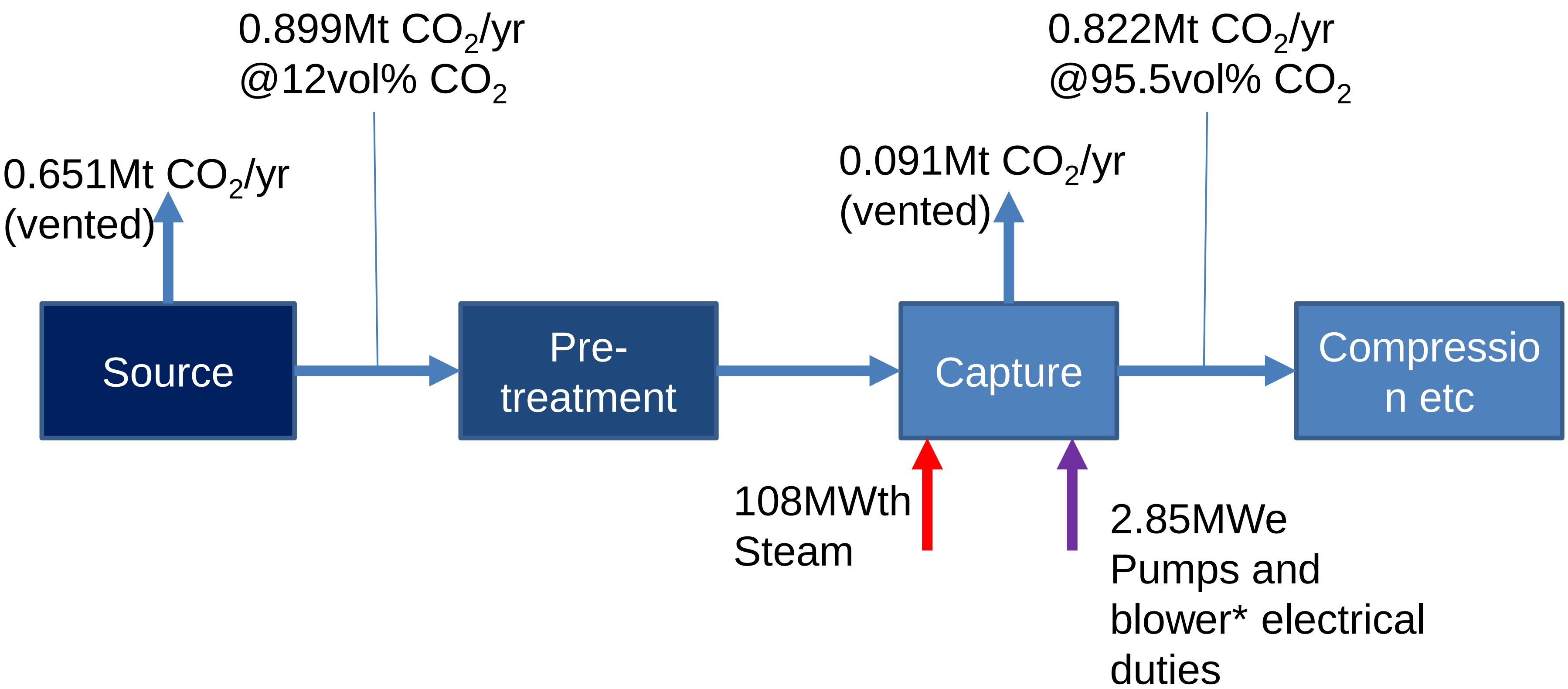
- The literature review, stakeholder interviews and high level techno-economic modelling highlight that, of technologies available for retrofit to refinery emissions today, implementation of 1st generation amine chemical solvent technologies could be attractive.
- A key uncertainty prioritised for examination, in agreement with stakeholders, through the process simulation is how the amount of $CO_2$ captured influences capture plant design.
- Therefore four sensitivities are considered involving different $CO_2$ streams and low and high load factors (described in the next slide).
- Source $CO_2$ stream data were kindly supplied by Ineos reflecting typical values at Grangemouth site.
- The four scenarios are named “part-load refinery (Baseline)”, “part-load refinery including CHP”, “pilot/demo” and “baseload refinery”.

# Source input assumptions (based on Ineos data)

| | Baseline Scenario #1 "Part-load refinery" | Sensitivity "Including CHP" (Scenario #2) | Sensitivity "Pilot/Demo" (Scenario #3) | Sensitivity "Baseload refinery" (Scenario #4) |
|---|---|---|---|---|
| Source $CO_2$ | 0.9 Mt/yr | 1.55 Mt/yr | 0.63Mt/yr | 1.55 Mt/yr |
| Cracker | Part Load | Part Load | Excluded | Full load |
| Other Refinery | Part Load | Part Load | Part Load | Full load |
| CHP | Excluded | Part Load | Excluded | Excluded |
| Average %$CO_2$ purity | 11.7 | 11.1 | 12.8 | 12.8 |

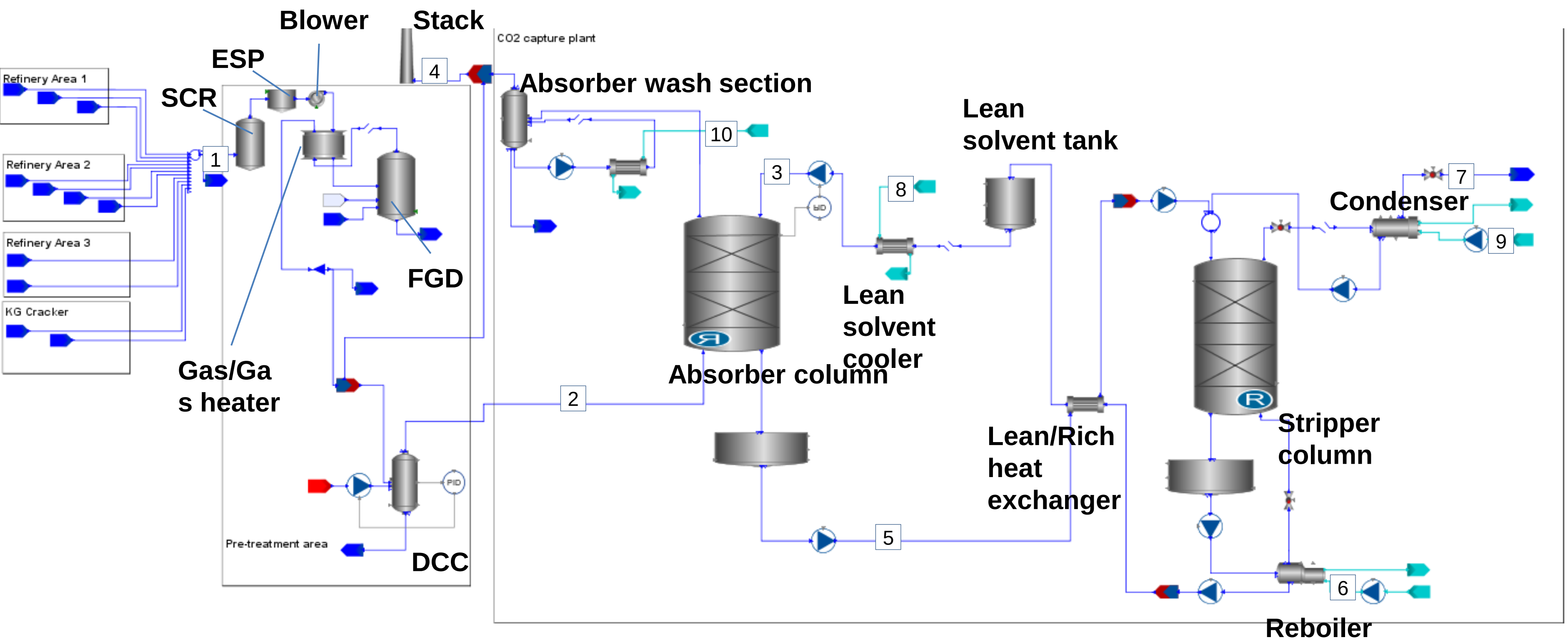
## Baseline Process Simulation

# High Level Process/Energy flows




* Blower is assumed to provide 0.05bar pressure difference

# Baseline Process Simulation




SCR – Selective Catalytic Reduction
ESP – Electrostatic Precipitator
FGD – Flue Gas Desulphurization
DCC – Direct Contact Cooler

# Stream Tables

| From: | | CO2 SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
|---|---|---|---|---|---|---|
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS | LEAN AMINE | OUTLET GAS | RICH AMINE |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 0 | 32,566 | 1,498,370 | 28,408 | 1,430,067 |
| **MEA** | | 0 | 0.0 | 702894.7 | 0.0 | 702480.6 |
| **$CO_2$** | | 107,229 | 109,765 | 141,718 | 10,877 | 239,642 |
| **$N_2$** | | 478,334 | 540,983 | 0 | 501,742 | 49 |
| **$O_2$** | | 33,441 | 0 | 0 | 38,904 | 0 |
| **$SO_2$** | | 3,977 | 0 | 0 | 41 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 27 | 0 | 0 | 27 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 624,420 | 683,314 | 2,342,982 | 580,217 | 2,372,239 |
| **Temperature** | $^oC$ | 266.8 | 41.0 | 40.8 | 40.0 | 51.0 |
| **Pressure** | bar(a) | 1.01 | 1.01 | 0.98 | 0.98 | 1.01 |

# Stream Tables (continued)

| From: | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
|---|---|---|---|---|---|---|
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 179,926 | 1,849 | 1,915,154 | 3,111,272 | 8,392,110 |
| **MEA** | | 0.0 | 0.9 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 97,911 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 50 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 179,926 | 99,811 | 1,915,154 | 3,111,272 | 8,392,110 |
| **Temperature** | ºC | 127.5 | 39.8 | 9.9 | 9.9 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
# Baseline scenario - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 1 |
| Source % $CO_2$ | 11.7 |
| Site total $CO_2$ capturable (tonnes/yr)* | 1,550,000 |
| % site $CO_2$ capturable | 58 |
| Reboiler Heat duty (MWth) | 108 |
| Total electrical power requirement of capture plant pumps (MWe) | 1.24 |
| Electrical power requirement of blower† (MWe) | 1.61 |
| Cooling water required (tonnes/hr) | 13413 |
| Capture plant site area required ($m^2$) | 14000 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 2300<br>SOx – 3000 |
| After (ppm) | NOx – 50<br>SOx – 6 |

* Based on Stream information provided by Ineos (for the refinery section alone)
† Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Baseline scenario – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 12.8 | Packing Height (m) | 18.4 | T/T Height (m) | 48.4 | 15,947,657 | 60.8 |
| Stripper | Diameter (m) | 7.9 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 3,868,691 | 14.8 |
| Reboiler | Heat Duty (MWth) | 107.8 | Steam flowrate (t/h) | 13.9 | | | 1,468,171 | 5.6 |
| Condenser | Cooling Duty (MWth) | 36.2 | Cooling water flowrate (t/h) | 240.1 | | | 505,037 | 1.9 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 130.9 | Heat transfer area per heat exchanger (m2) | 436.9 | Number of heat exchangers | 25 | 1,448,633 | 5.5 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 435,924 | 1.7 |
| Lean amine cooler | Cooling Duty (MWth) | 44.5 | Heat transfer area per heat exchanger (m2) | 665.0 | Number of heat exchangers | 4 | 667,908 | 2.5 |
| Rich solvent pump | Total power requirement (kW) | 648.5 | Number of pumps required | 666.5 | | | 187,312 | 0.7 |
| Lean solvent pump | Total power requirement (kW) | 244.6 | Number of pumps required | 658.3 | | | 92,956 | 0.4 |
| Cooling water pumps | Total power requirement (kW) | 354.8 | Number of pumps required | 1395.6 | | | 385,311 | 1.5 |
| Steam boiler | Capacity (t/h steam) | 179.9 | | | | | 1,218,264 | 4.6 |
| Total equipment purchase cost (PCE) | | | | | | | **26,225,864** | |

# Simulation Results
# Baseline scenario - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 15,947,657 | 60.8 |
| Stripper | 3,868,691 | 14.8 |
| Reboiler | 1,468,171 | 5.6 |
| Condenser | 505,037 | 1.9 |
| Lean/Rich Heat Exchanger | 1,448,633 | 5.5 |
| Lean amine tank | 435,924 | 1.7 |
| Lean amine cooler | 667,908 | 2.5 |
| Rich solvent pump | 187,312 | 0.7 |
| Lean solvent pump | 92,956 | 0.4 |
| Cooling water pumps | 385,311 | 1.5 |
| Steam boiler | 1,218,264 | 4.6 |
| **Total equipment purchase cost (PCE)** | **26,225,864** | |

# Simulation Results

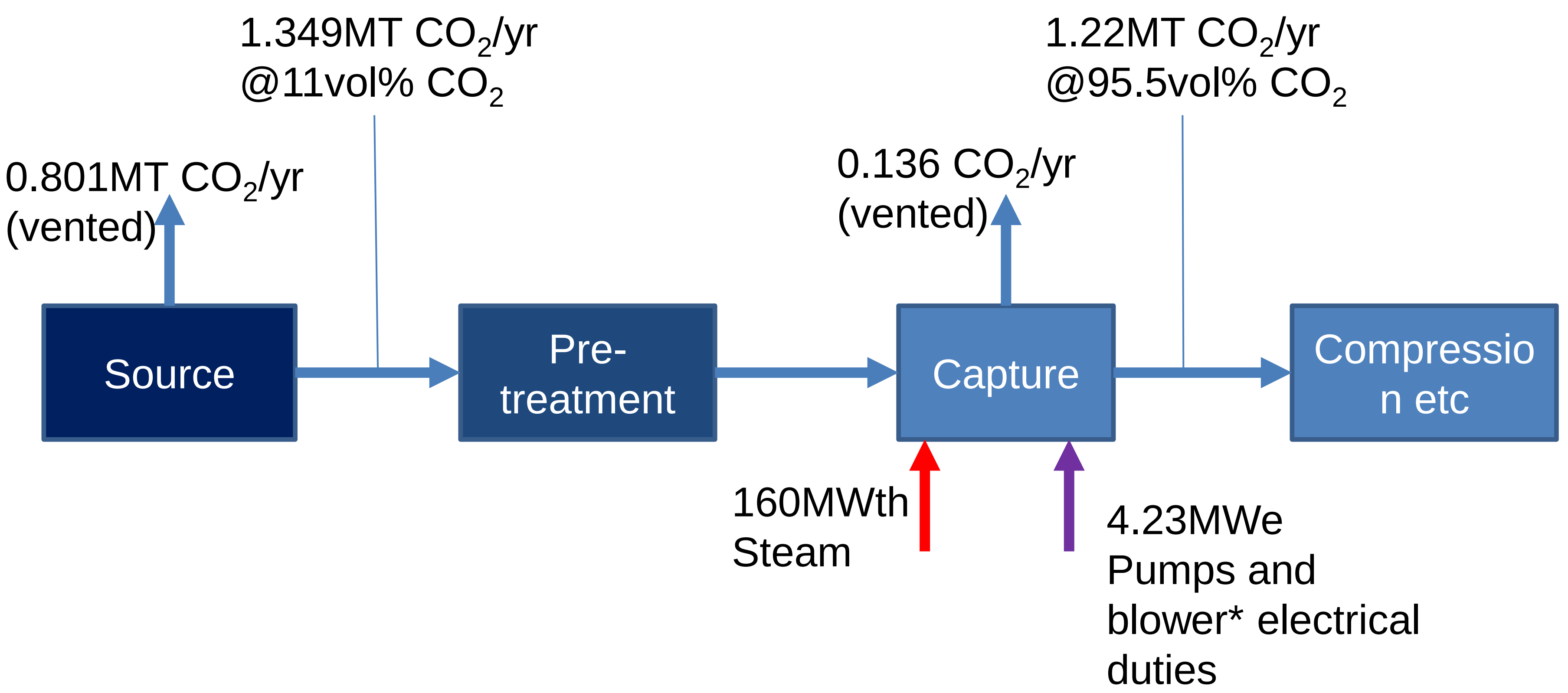
## Baseline scenario - Capital Expenditure is 1/6th of the total fixed capital cost.

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 26,225,864 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 83,922,766 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **166,418,845** |

# Simulation Results
## Baseline scenario - Operating Expenditure

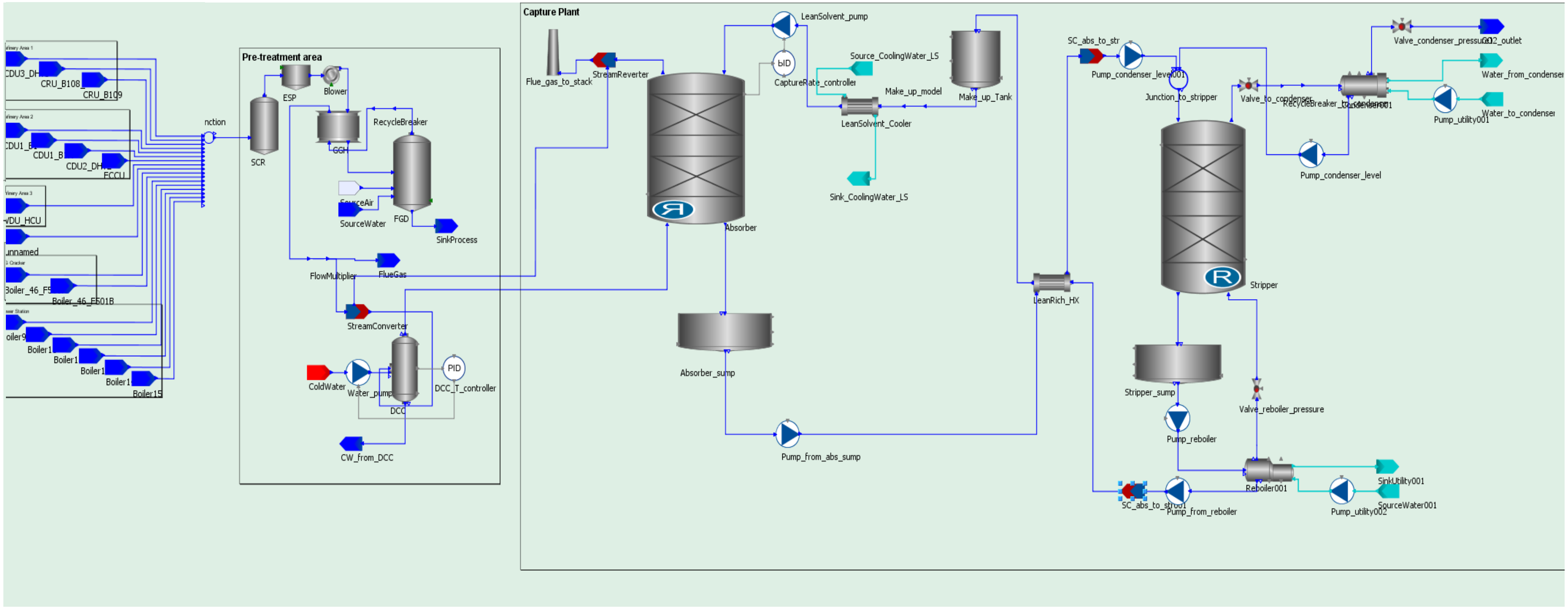
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 8,320,942 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 832,094 |
| Solvent make-up cost | | 4,150,461 |
| Pumps power cost | | 793,570 |
| Utilities - Steam costs | | 21,494,013 |
| Utilities - Cooling water costs | | 633,059 |
| **Total Variable costs** | | **27,903,198** |
| **OPEX** | | **36,224,140** |

# Scenario 2 Process Simulation
# High Level Process/Energy flows (including power plant)




* Blower is assumed to provide 0.05bar pressure difference

# Scenario 2 Process Simulation
# All $CO_2$ emissions (including power plant)



# Simulation Results
# Scenario 2 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 1 |
| Source vol % CO2 | 11.1 |
| Site total CO2 captureable* (tonnes/year) | 1550000 |
| % site CO2 captureable | 87 |
| Total reboiler heat duty (MWth) | 160 |
| Reboiler Specific duty (GJ/t CO2) | 3.96 |
| Total electrical power requirement of capture plant pumps (MWe) | 1.85 |
| Electrical power requirement of blower† (MWe) | 2.38 |
| Total Cooling water required (tonnes/hr) | 20332 |
| Total Capture plant site area required (m2) | 21000 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.08 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 2300<br>SOx – 2300 |
| After (ppm) | NOx – 50<br>SOx – 6 |

* Based on Stream information provided by Ineos (for the refinery section alone)
† Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 2 – Equipment List

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 15.7 | Packing Height (m) | 18.2 | T/T Height (m) | 48.2 | 23,060,176 | 59 |
| Stripper | Diameter (m) | 9.6 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 6,290,073 | 16 |
| Reboiler | Heat Duty (MWth) | 160.5 | Steam flowrate (t/h) | 20.7 | | | 2,260,065 | 6 |
| Condenser | Cooling Duty (MWth) | 53.9 | Cooling water flowrate (t/h) | 358.3 | | | 1,194,010 | 3 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 197.0 | Heat transfer area per heat exchanger (m2) | 653.7 | Number of heat exchangers | 25 | 2,076,362 | 5 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 542,357 | 1 |
| Lean amine cooler | Cooling Duty (MWth) | 64.3 | Heat transfer area per heat exchanger (m2) | 986.7 | Number of heat exchangers | 4 | 1,008,586 | 3 |
| Rich solvent pump | Total power requirement (kW) | 964.8 | Number of pumps required | 16.0 | | | 267,311 | 1 |
| Lean solvent pump | Total power requirement (kW) | 364.1 | Number of pumps required | 8.0 | | | 132,695 | 0 |
| Cooling water pumps | Total power requirement (kW) | 524.0 | Number of pumps required | 34.0 | | | 559,560 | 1 |
| Steam boiler | Capacity (t/h steam) | 267.7 | | | | | 1,774,210 | 5 |
| Total equipment purchase cost (PCE) | | | | | | | 39,165,405 | |

# Simulation Results
# Scenario 2 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 23,060,176 | 58.9 |
| Stripper | 6,290,073 | 16.1 |
| Reboiler | 2,260,065 | 5.8 |
| Condenser | 1,194,010 | 3.0 |
| Lean/Rich Heat Exchanger | 2,076,362 | 5.3 |
| Lean amine tank | 542,357 | 1.4 |
| Lean amine cooler | 1,008,586 | 2.6 |
| Rich solvent pump | 267,311 | 0.7 |
| Lean solvent pump | 132,695 | 0.3 |
| Cooling water pumps | 559,560 | 1.4 |
| Steam boiler | 1,774,210 | 4.5 |
| **Total equipment purchase cost (PCE)** | **39,165,405** | |

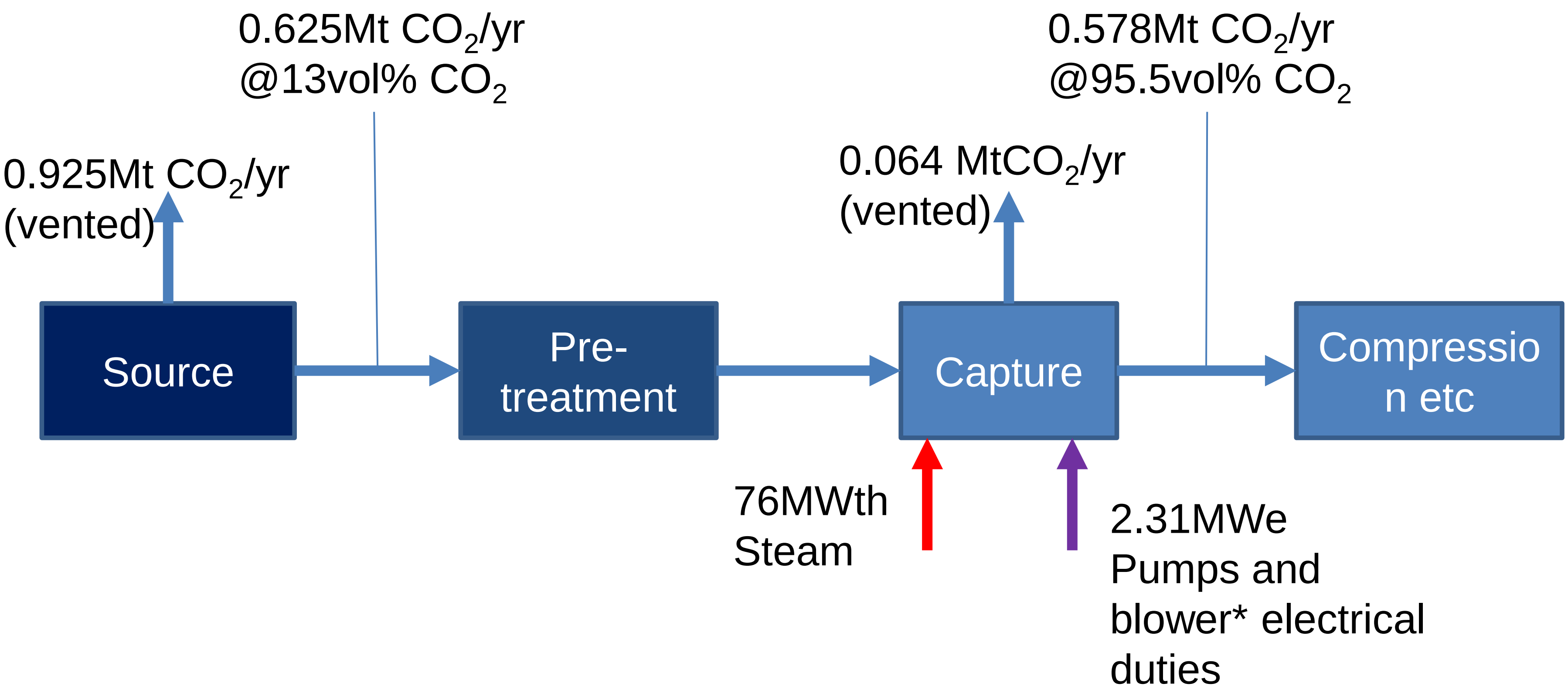
# Simulation Results
# Scenario 2 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 39,165,405 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 125,329,297 |
| Contigency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **248,527,996** |

# Simulation Results
# Scenario 2 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 12,426,400 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 1,242,640 |
| Solvent make-up cost | | 6,168,646 |
| Pumps power cost | | 1,178,253 |
| Utilities - Steam costs | | 31,978,574 |
| Utilities - Cooling water costs | | 933,458 |
| **Total Variable costs** | | **41,501,571** |
| **OPEX** | | **53,927,971** |

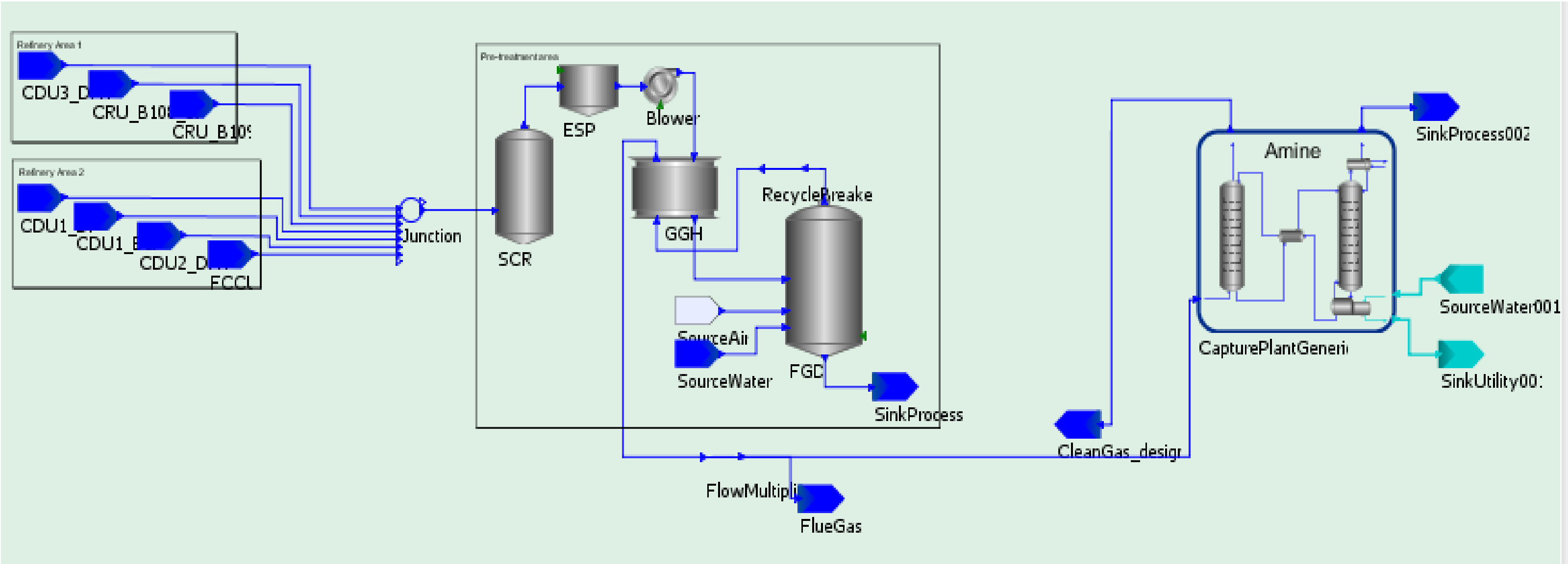
## Scenario 3 Process Simulation
## High Level Process/Energy flows (Refinery Areas 1 and 2)




* Blower is assumed to provide 0.05bar pressure difference

# Scenario 3 Process Simulation
# Refinery Areas 1 and 2

- Scenario carried out using the simplified capture plant.
- Assumed values highlighted in red (assumed equal to Baseline scenario)



# Simulation Results
# Scenario 3 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 1 |
| Source % $CO_2$ | 12.8 |
| Site total $CO_2$ capturable (tonnes/year)* | 1550000 |
| % site $CO_2$ capturable | 40.5 |
| Reboiler Heat duty (MWth) | 75.9 |
| Total electrical power requirement of capture plant pumps (MWe) | 3.96 |
| Electrical power requirement of blower† (MWe) | 1.24 |
| Cooling water required (tonnes/hr) | 1.07 |
| Capture plant site area required ($m^2$) | 13413 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 2300<br>SOx – 3500 |
| After (ppm) | NOx – 50<br>SOx – 6 |

* Based on Stream information provided by Ineos (for the refinery section alone)
† Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 3 – Equipment List

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 10.5 | Packing Height (m) | 18.4 | T/T Height (m) | 48.4 | 11,158,190 | 57 |
| Stripper | Diameter (m) | 6.7 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 2,900,399 | 15 |
| Reboiler | Heat Duty (MWth) | 107.8 | Steam flowrate (t/h) | 13.9 | | | 1,004,933 | 5 |
| Condenser | Cooling Duty (MWth) | 36.2 | Cooling water flowrate (t/h) | 239.9 | | | 555,636 | 3 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 130.7 | Heat transfer area per heat exchanger (m2) | 436.6 | Number of heat exchangers | 25 | 1,447,526 | 7 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 360,059 | 2 |
| Lean amine cooler | Cooling Duty (MWth) | 44.5 | Heat transfer area per heat exchanger (m2) | 663.5 | Number of heat exchangers | 4 | 737,068 | 4 |
| Rich solvent pump | Total power requirement (kW) | 648.1 | Number of pumps required | 12.0 | | | 187,238 | 1 |
| Lean solvent pump | Total power requirement (kW) | 244.5 | Number of pumps required | 4.0 | | | 64,045 | 0 |
| Cooling water pumps | Total power requirement (kW) | 354.1 | Number of pumps required | 24.0 | | | 385,265 | 2 |
| Steam boiler | Capacity (t/h steam) | 124.7 | | | | | 868,411 | 4 |
| Total equipment purchase cost (PCE) | | | | | | | 19,668,771 | |

# Simulation Results
# Scenario 3 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 11,158,190 | 56.7 |
| Stripper | 2,900,399 | 14.7 |
| Reboiler | 1,004,933 | 5.1 |
| Condenser | 555,636 | 2.8 |
| Lean/Rich Heat Exchanger | 1,447,526 | 7.4 |
| Lean amine tank | 360,059 | 1.8 |
| Lean amine cooler | 737,068 | 3.7 |
| Rich solvent pump | 187,238 | 1.0 |
| Lean solvent pump | 64,045 | 0.3 |
| Cooling water pumps | 385,265 | 2.0 |
| Steam boiler | 868,411 | 4.4 |
| **Total equipment purchase cost (PCE)** | **19,668,771** | |

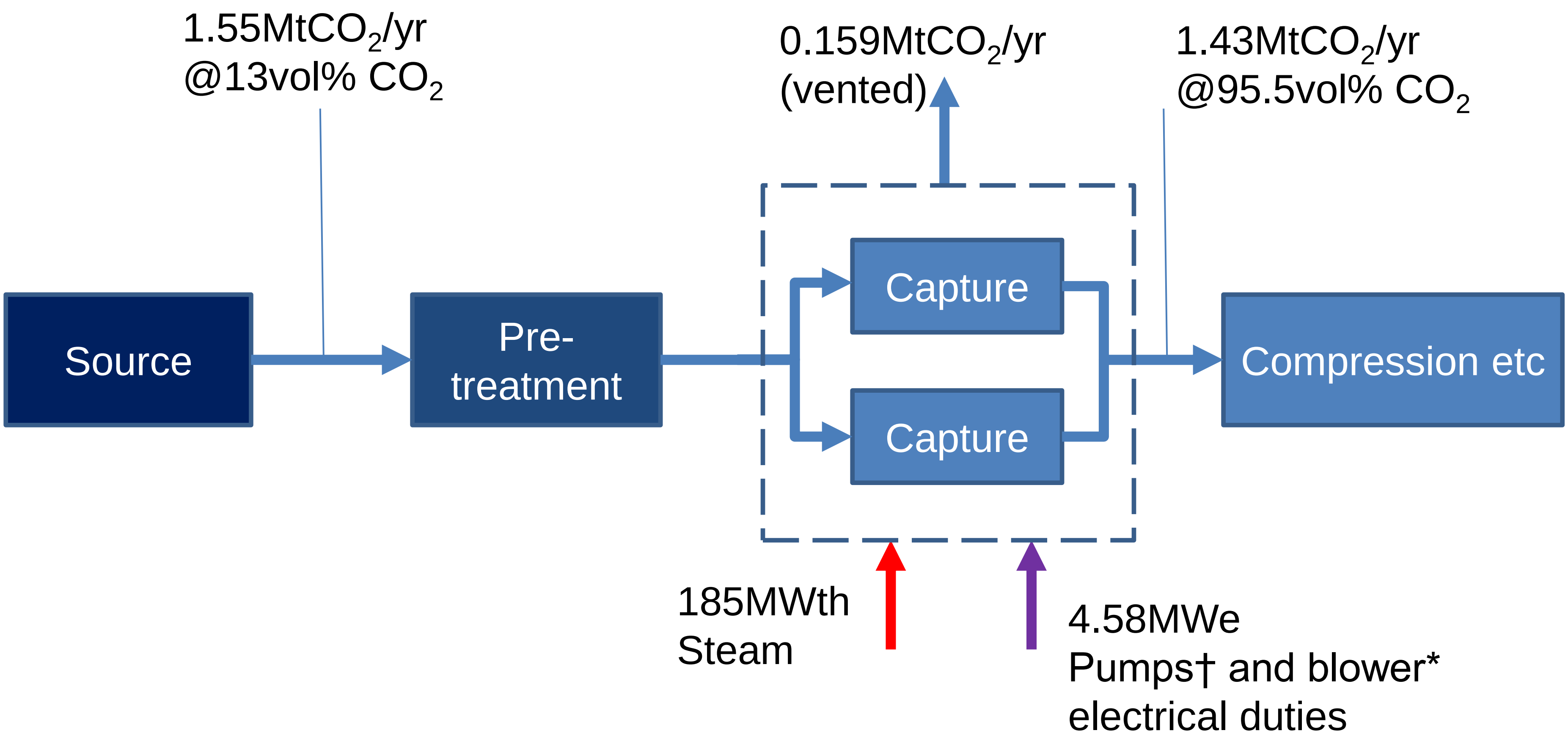
# Simulation Results
# Scenario 3 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 19,668,771 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 62,940,067 |
| Contigency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **124,810,153** |

# Simulation Results
# Scenario 3 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 6,240,508 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 624,051 |
| Solvent make-up cost | | 4,148,325 |
| Pumps power cost | | 766,584 |
| Utilities - Steam costs | | 14,812,356 |
| Utilities - Cooling water costs | | 632,937 |
| **Total Variable costs** | | **20,984,253** |
| **OPEX** | | **27,224,761** |

# Scenario 4 Process Simulation
# High Level Process/Energy flows (Refinery Areas 1, 2 and 3 full flow)

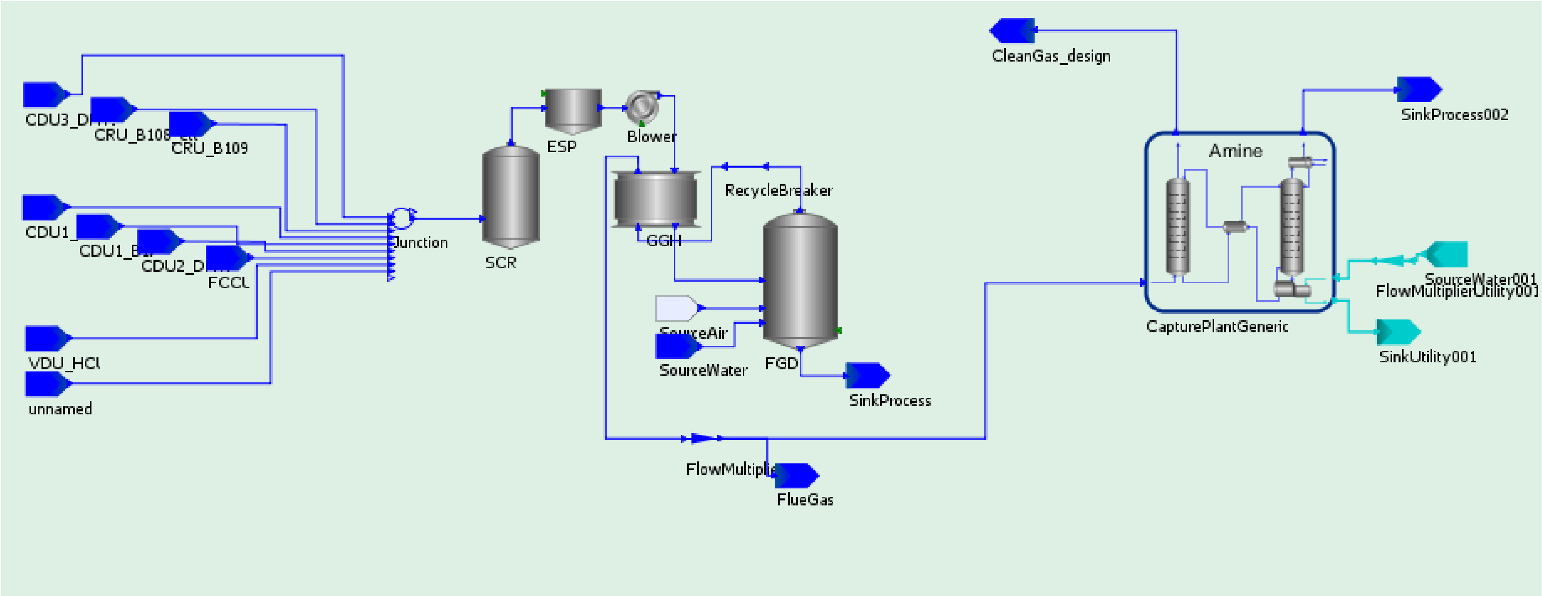



* Blower is assumed to provide 0.05bar pressure difference

† Pump power assumed to be same as baseline

# Scenario 4 Process Simulation
# Refinery Areas 1, 2 and 3 full flow

- Scenario carried out using the simplified capture plant. All flows are scaled up to match emission levels (1.55Mt $CO_2$/year)
- Assumed values highlighted in red (assumed equal to Baseline scenario)



# Simulation Results
# Scenario 4 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 2 |
| Source % $CO_2$ | 12.8 |
| Site total $CO_2$ captureable (tonnes/year) | 1550000 |
| % site $CO_2$ captureable | 100 |
| Total Reboiler Heat duty (MWth) | 185 |
| Reboiler Specific duty (GJ/t CO2) | 3.97 |
| Total electrical power requirement of capture plant pumps (MWe) | 1.85 |
| Electrical power requirement of blower† (MWe) | 2.73 |
| Capture plant site area required ($m^2$) | 9500 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 2300<br>SOx – 3500 |
| After (ppm) | NOx – 50<br>SOx – 6 |

* Based on Stream information provided by Ineos (for the refinery section alone)
† Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 4 – Equipment List (for 1 train of capture plant)

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 11.8 | Packing Height (m) | 18.4 | T/T Height (m) | 48.4 | 13,505,813 | 58.7 |
| Stripper | Diameter (m) | 7.4 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 3,501,938 | 15.2 |
| Reboiler | Heat Duty (MWth) | 107.8 | Steam flowrate (t/h) | 13.9 | | | 1,266,927 | 5.5 |
| Condenser | Cooling Duty (MWth) | 36.2 | Cooling water flowrate (t/h) | 239.9 | | | 504,691 | 2.2 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 130.7 | Heat transfer area per heat exchanger (m2) | 436.6 | Number of heat exchangers | 25 | 1,425,459 | 6.2 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 404,739 | 1.8 |
| Lean amine cooler | Cooling Duty (MWth) | 44.5 | Heat transfer area per heat exchanger (m2) | 663.5 | Number of heat exchangers | 4 | 666,454 | 2.9 |
| Rich solvent pump | Total power requirement (kW) | 648.1 | Number of pumps required | 354035.3 | | | 187,238 | 0.8 |
| Lean solvent pump | Total power requirement (kW) | 244.5 | Number of pumps required | 87845.5 | | | 92,917 | 0.4 |
| Cooling water pumps | Total power requirement (kW) | 354.1 | Number of pumps required | 728470.5 | | | 385,265 | 1.7 |
| Steam boiler | Capacity (t/h steam) | 154.7 | | | | | 1,058,265 | 4.6 |
| Total equipment purchase cost (PCE) | | | | | | | 22,999,707 | |

# Simulation Results
# Scenario 4 - Capital Expenditure (per train)

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 27,011,626 | 58.7 |
| Stripper | 7,003,877 | 15.2 |
| Reboiler | 2,533,853 | 5.5 |
| Condenser | 1,009,382 | 2.2 |
| Lean/Rich Heat Exchanger | 2,850,918 | 6.2 |
| Lean amine tank | 809,478 | 1.8 |
| Lean amine cooler | 1,332,909 | 2.9 |
| Rich solvent pump | 374,477 | 0.8 |
| Lean solvent pump | 185,835 | 0.4 |
| Cooling water pumps | 770,531 | 1.7 |
| Steam boiler | 2,116,529 | 4.6 |
| **Total equipment purchase cost (PCE)** | **45,999,413** | |

# Simulation Results
# Scenario 4 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 45,999,413 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 147,198,122 |
| Contigency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **291,893,876** |

# Simulation Results
# Scenario 4 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 14,594,694 |
| **Variable Costs** | | |
| Solvent make-up costs | Based on the amount of $CO_2$ captured | 8,296,651 |
| Miscellaneous materials cost | 10% of maintenance costs | 1,459,469 |
| Pumps power cost | | 1,585,531 |
| Utilities - Steam costs | | 36,953,196 |
| Utilities - Cooling water costs | | 1,265,874 |
| **Total Variable costs** | | **49,560,721** |
| **OPEX** | | **64,155,415** |

# Comparison of costs between scenarios

| | Baseline Scenario #1 "Part-load refinery" | Sensitivity "Including CHP" (Scenario #2) | Sensitivity "Pilot/Demo" (Scenario #3) | Sensitivity "Baseload refinery" (Scenario #4) |
|---|---|---|---|---|
| Source $CO_2$ | 0.90 Mt/yr | 1.35 Mt/yr | 0.63Mt/yr | 1.55 Mt/yr |
| Equipment cost | £26m | £39m | £20m | £46m |
| Total fixed cost | £140m | £209m | £105m | £246m |
| Annual opex (incl. energy) | £36m/yr | £54m/yr | £27m/yr | £64m/yr |
| Reboiler Heat Duty $MW_{th}$ | 108 | 160 | 76 | 185 |
| Power /$MW_e$ | 2.85 | 4.23 | 2.31 | 4.58 |

# Insights from process simulation

- Capture with a 1st generation amine solvent at a UK refinery is feasible, but there are many potential configurations that should be considered, primarily which source streams should be captured, and how much of these.
- The overal capital cost is more than five times the cost of the main pieces of equipment.
- The largest of the equipment cost items modelled is the absorber.
- The largest operating cost modelled is for steam.
- Therefore capture technology development should focus on reducing the costs of absorber, and/or the amount of steam required, and simplifying retrofit installation.
- The feasibility, design and costing of pre-treatment including scrubbing equipment and a network for gathering and managing $CO_2$ flows together from diverse sources on a refinery will require significant site specific analysis (not possible in this study).
- For reasons of operability, reliability, flexibility, as well as commercial availability of equipment, installations capturing above *ca*. 1-1.4 $MtCO_2$/yr may adopt configurations involving two absorber trains rather than one.
- Water requirements are feasible but will require relevant permits from the Environment Agency.

# The outputs from the process simulation can be inform future techno-economic studies.

| Parameter | Techno-economics “Baseline” | Process simulation “Baseline” |
|---|---|---|
| Input flue gas $MtCO_2$/yr | 1.6 Mt/yr | 0.9 Mt/yr |
| Abated $MtCO_2$/yr | 0.7 $MtCO_2$/yr | Not calculated directly |
| Capex | Capture only: £281m<br>(£489m incl. pre-treatment) | Capture only: £166m |
| Non-energy opex | £18m/yr | £14m/yr |
| Heat | 109 MW | 108 MW |
| Power | 6.5 MW | 2.9 MW |

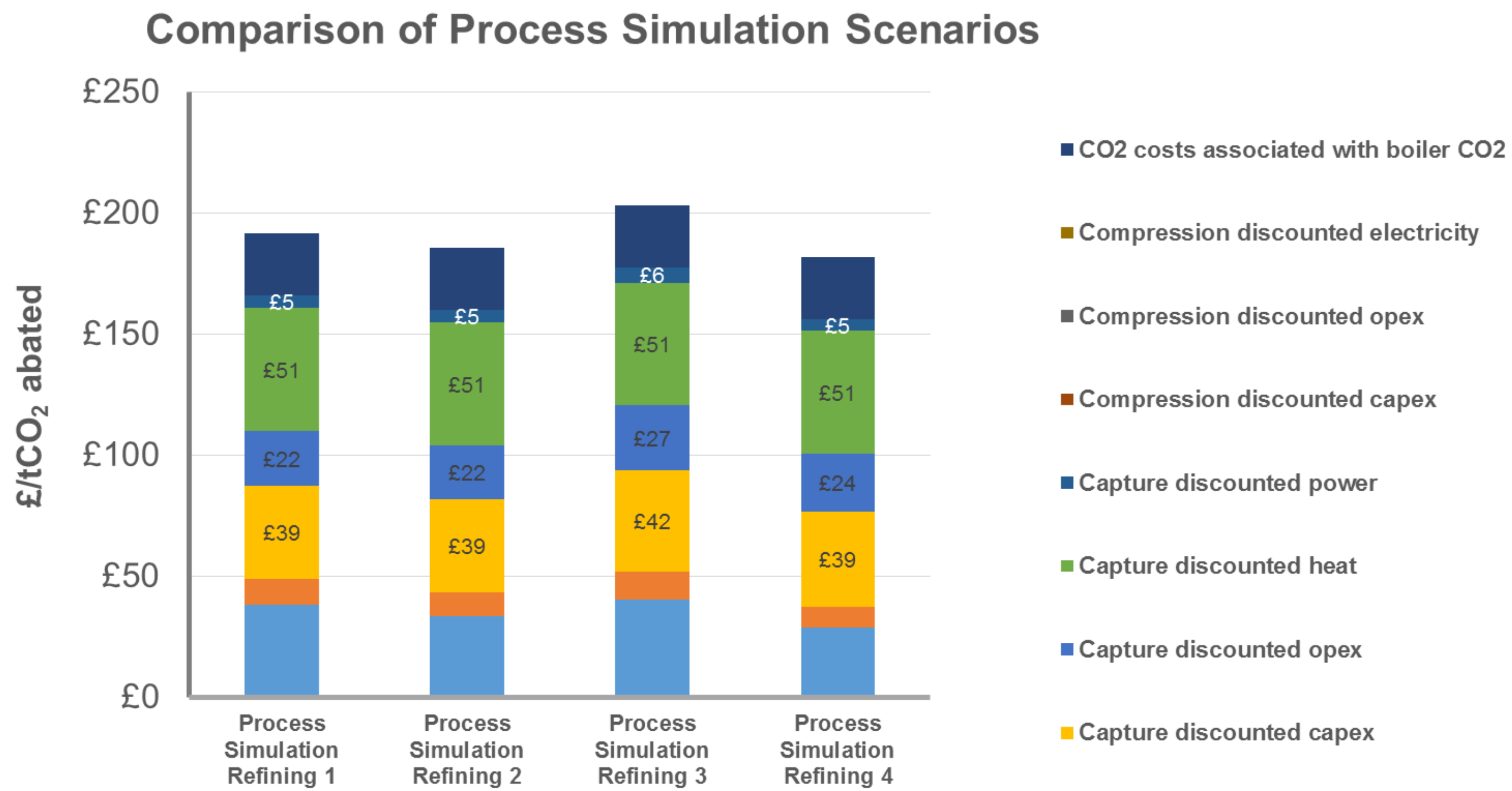
# Issues emerging from process simulation of MEA-refinery configurations

- It is not surprising that opposite approaches to estimating costs of capture do not agree completely. Cost similarities for techno-economics and process simulation are being examined, although analysis may be limited by poor description in published reports and limited realised UK experience of constructing similar plants.
- If a policy objective is to implement close to full scale capture (i.e. 1-3 Mt/yr) at a UK refinery in the 2020s, then the key barrier to overcome is experience of capture of refinery gases at an appropriate scale.
- A plausible development strategy to overcome this barrier could be to begin with a pilot/demo project capturing ca. 0.6 $MtCO_2$/yr by 2020 using a single train 1$^{st}$ generation amine system. This pilot/demo would be similar to the Source #20 in the techno-economic model or Scenario #3 in the process simulation.
- An initial project would likely draw on a single vent type (for simplicity) and single train amine plant.
- Following successful experience with this, it should be possible to either ramp up capture capacity by 2025 to capture 1.5-2.5 Mt/yr (by bringing together multiple streams with multiple trains) or reduce unit costs by employing second generation amines or blends, likely using a separate amine train.
- Engineering studies are time consuming  so need to begin early, and will be required for each site to understand the optimum infrastructure to bring together multiple, diverse $CO_2$ streams.

# Comparison of levelised costs between scenarios

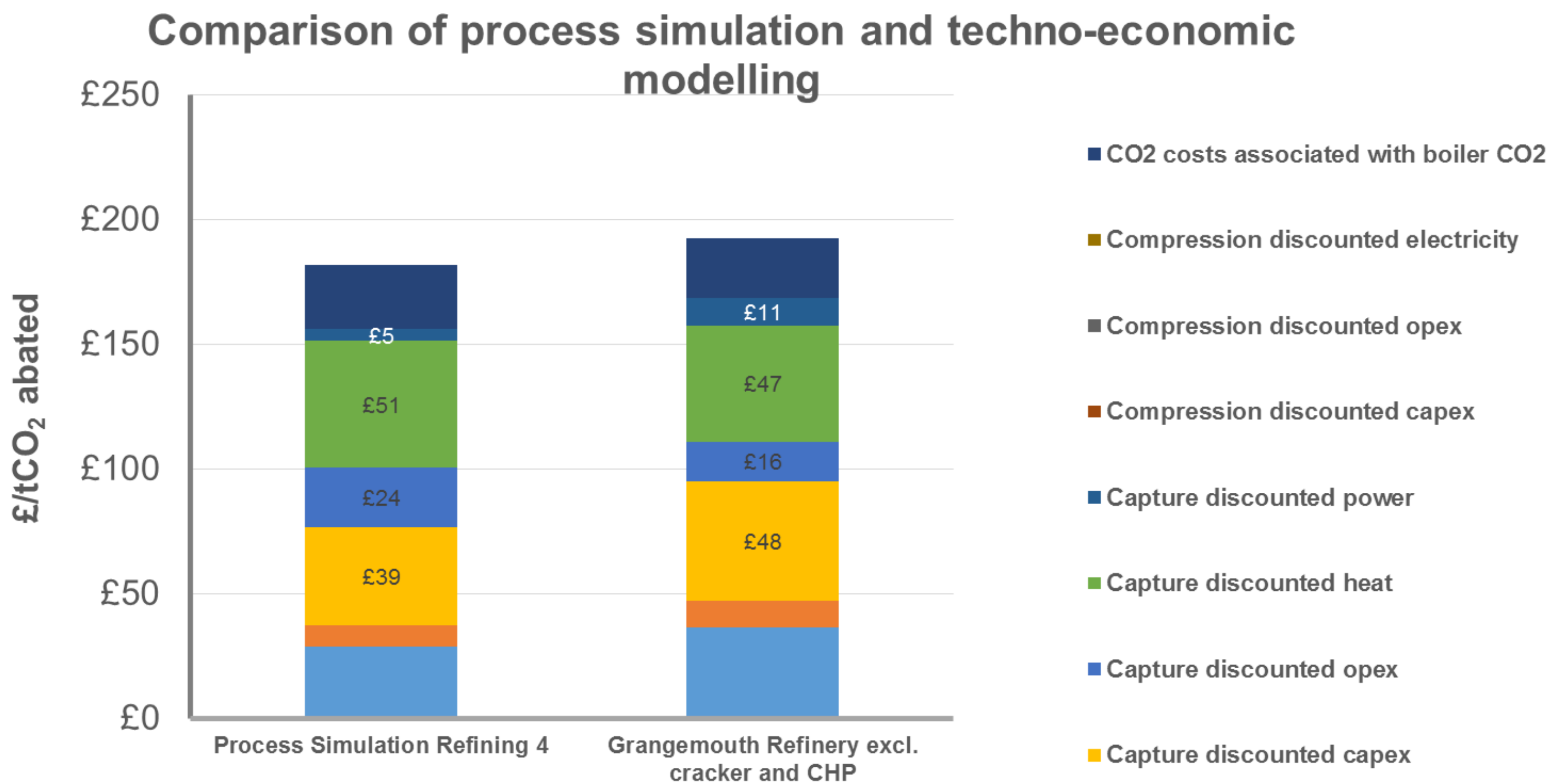
- The primary aim for the process simulation is to understand design requirements and order of magnitude cost drivers (rather than absolute costs).
- To facilitate cost comparison between scenarios, overall levelised costs have been calculated by including the data on capacity, capex, fixed opex, heat and power requirements emerging from the process simulation in the techno-economic model. The pre-treatment, energy/carbon prices and structure of the discounted cashflow is otherwise identical to that used in the Pragmatic scenario.
- Note that the process designs are illustrative and not "optimised" (optimisation of process simulations is highly resource intensive and out of the scope of the present study). Therefore small differences in sizes, energy requirements and costs between scenarios should not be over-interpreted.

# Similar levelised costs for the different process simulations




| Process Simulation Refining 1 | Process Simulation Refining 2 | Process Simulation Refining 3 | Process Simulation Refining 4 |
|---|---|---|---|
| Baseline Refinery part-load 11.7% $CO_2$ 0.9Mt/yr input 0.62 Mt/yr abated | Baseline + CHP 11.1% $CO_2$ 1.55 Mt/yr input 0.94 Mt/yr abated | Pilot/Demo 12.8% $CO_2$ 0.64 Mt/yr input 0.43 Mt/yr abated | Refinery baseload 12.8% $CO_2$ 1.55 Mt/yr input 1.0 Mt/yr abated |

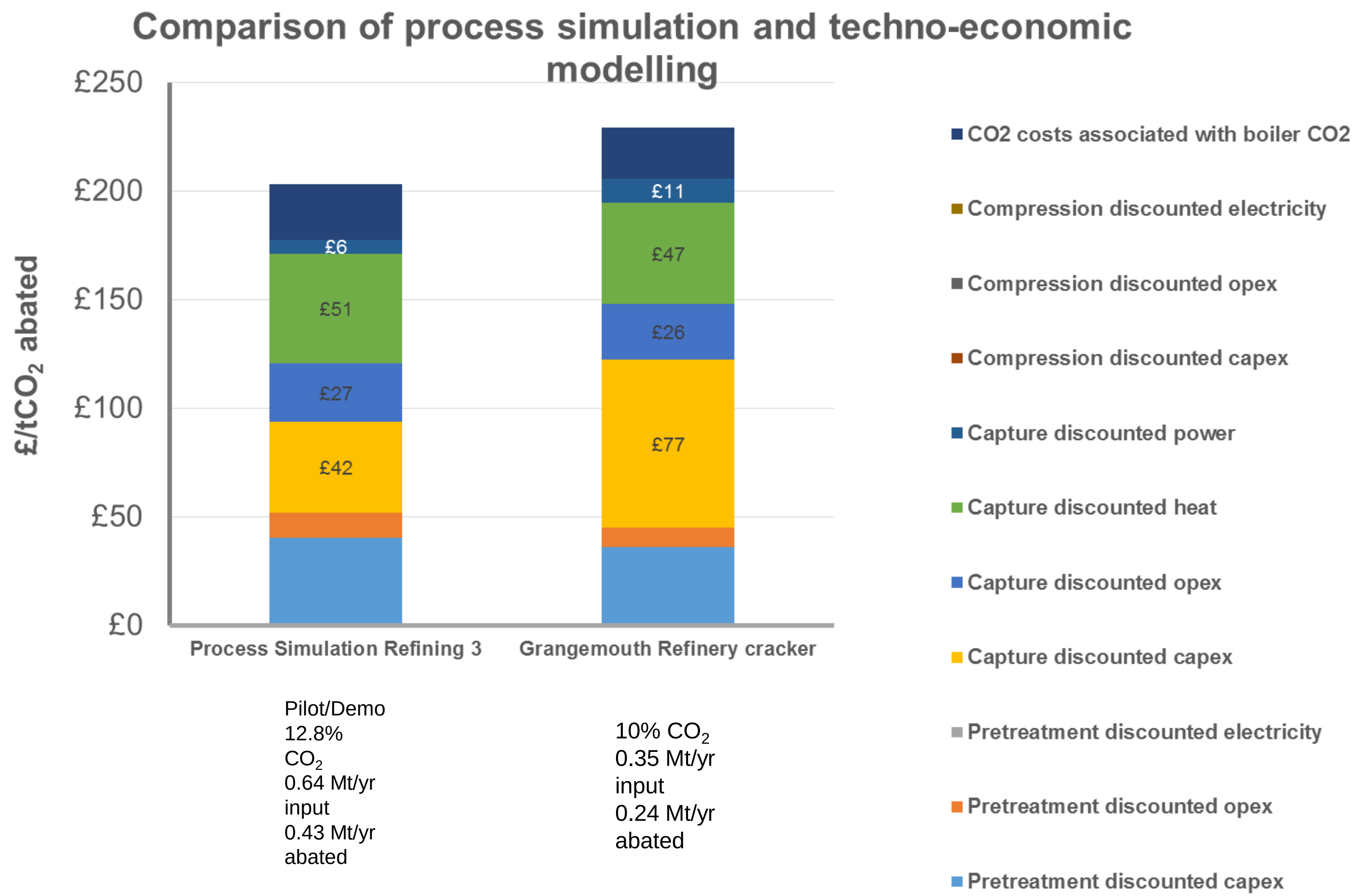
# Abatement cost breakdown for medium size refinery project




Refinery baseload 12.8% $CO_2$ 1.55 Mt/yr input 1.0 Mt/yr abated

10% $CO_2$ Site 1.3 Mt/yr captured out of 1.62 Mt/yr input 1.0 Mt/yr abated

# Abatement cost breakdown for a “cracker only” project




Pilot/Demo
12.8% $CO_2$
0.64 Mt/yr input
0.43 Mt/yr abated

10% $CO_2$
0.35 Mt/yr input
0.24 Mt/yr abated

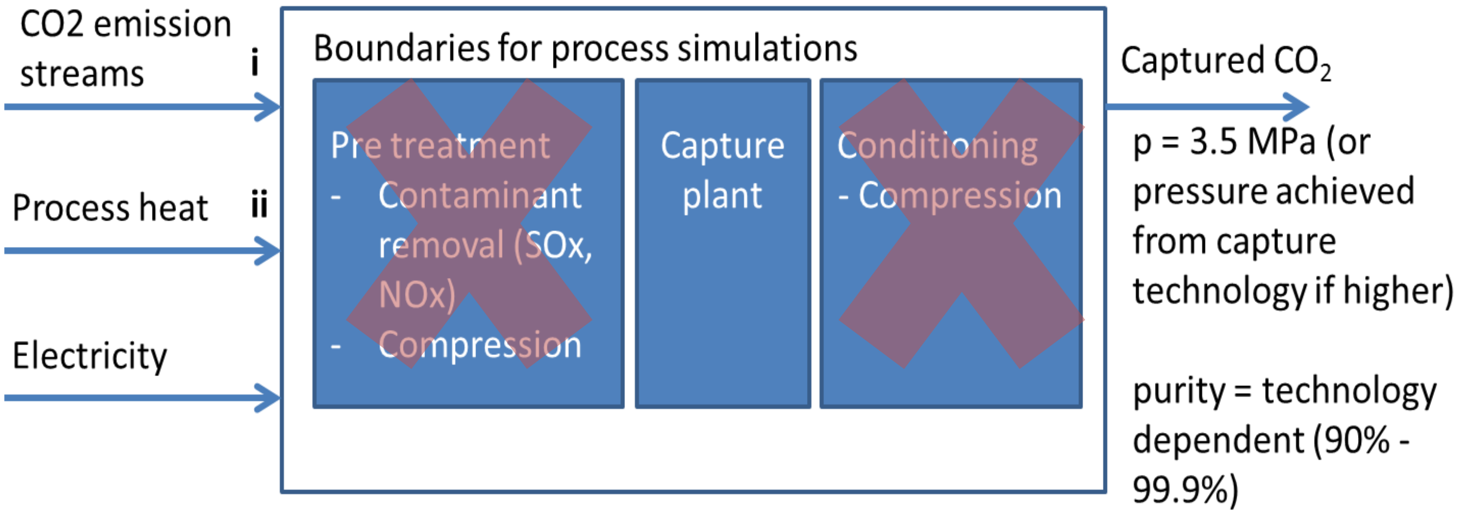
# Four case studies on process simulation

1. Grangemouth oil refinery with MEA capture
2. Lafarge Tarmac Dunbar cement works with MEA capture
3. Tata Scunthorpe iron and steel plant with MEA capture at CHP unit
4. GrowHow ammonia production site with $CO_2$ compression

# Process Simulation

- The literature review, stakeholder interviews, and techno-economic modelling confirmed that 1st generation amine technologies could be attractive capture technologies for any UK industrial retrofit in the period 2020 to 2025 (among other technologies).
- Key uncertainties prioritised from process simulation are the impacts of $CO_2$ concentration and project scale – these are reflected in the choice of three scenarios for process simulation (named “1-Baseline”, “2-Pilot” and “3-High $CO_2$ Concentration”).
- The Lafarge Tarmac Dunbar and the Heidelberg Ribblesdale cement works kindly provided typical $CO_2$ stream conditions used as inputs for the process simulation.

# Scope of study




**i** - only process emissions (excluding power and heat production emissions)
- in case multiple stack emissions are combined, piping etc to bring together is included

**ii** - including integration cost in case of process heat recovery
- assume natural gas boiler if no process heat is available (default assumption)

Further Notes

- Compression plant not simulated
- Pre-treatment processes simulated but not costed

# Inputs

- 30 wt% MEA solvent
- 90% capture target
- >95% purity $CO_2$ by volume
- Absorber operating at atmospheric pressure
- Mellapak 250Y structured packing used in Absorber and stripper columns
- Heat exchanger overall heat transfer coefficient – 6000W/m$^2$K
- Stripper feed temperature - 102°C
- Maximum flow in heat exchanger – 2500m$^3$/hr
- Stripper operating pressure ~ 1.67bara
- Steam pressure 3.5bara
- Cooling water temperature is assumed to be 10°C
- 8400 hours of continuous operation assumed in a year

# Detailed assumptions

- Requirements to convey flue sources to capture plant are ignored
- A basic MEA $CO_2$ capture plant configuration is selected for the study (complex configurations such as split-flow considerations are ignored)
- An optimal lean loading of 0.28 mol $CO_2$/mol MEA was assumed for the scenarios. This may introduce some inefficiencies in the performance. A full optimization for each scenario would require substantial development work.
- The absorber packing height was sized to achieve a rich loading of 0.473 mol $CO_2$/mol MEA. The rich loading value used was assumed to be representative of the physical constraint on the capacity of the MEA solvent in chemical absorption processes.
- The column models were rate-based, distributed models. A trade-off between simulation performance (times) and accuracy was made reducing the number of discretization elements. This resulted in slight mass imbalances for certain components of up to about 1%.
- $O_2$/$N_2$ composition in flue gas streams are lumped as inert material
- $SO_X$ and $NO_X$ compositions are not considered at capture plant boundary
- Solvent degradation effects ignored
- As degradation rates are not estimated, no reclaimer is modelled or costed in this unit.

# Detailed assumptions (continued)

- The site area required for the capture plant is based on published work[1] and scaled as linearly dependent on the absorber diameter. This was done based on the fact that the absorber typically presents the largest footprint on the plant.
- In the pre-treatment area, the blower was assumed to provide a pressure increase of 0.05bar to overcome pressure drops in the downstream units.
- $CO_2$ emissions from the standalone steam boiler are not considered in this study.
- Cooling water requirements for the absorber wash water section are rough estimates. A trade-off is required between how much (and under what conditions) wash water is supplied to the wash section and the required cooling water flowrate. The optimisation of this scheme was outside the scope of this study and as a result was not carried out. A fixed flowrate of wash water was assumed and the cooling water flowrate to cool that stream is estimated.
- It should be noted that the same “hand factors” were used to estimate the total fixed capital costs from the total equipment purchase cost for all scales of the capture plants modelled. It may be more appropriate to employ different factors for different scales of plants.
- Fuel and electricity prices used are based on DECC’s 2025 costs for all the demo scenarios and 2020 cost for the pilot plant scenario.
- All other costs are based on Q3 2013. These could be translated to the same basis by assuming an appropriate discount rate.

[1]IEAGHG (2012). $CO_2$ capture at gas-fired power plants, Report No: 2012/8.

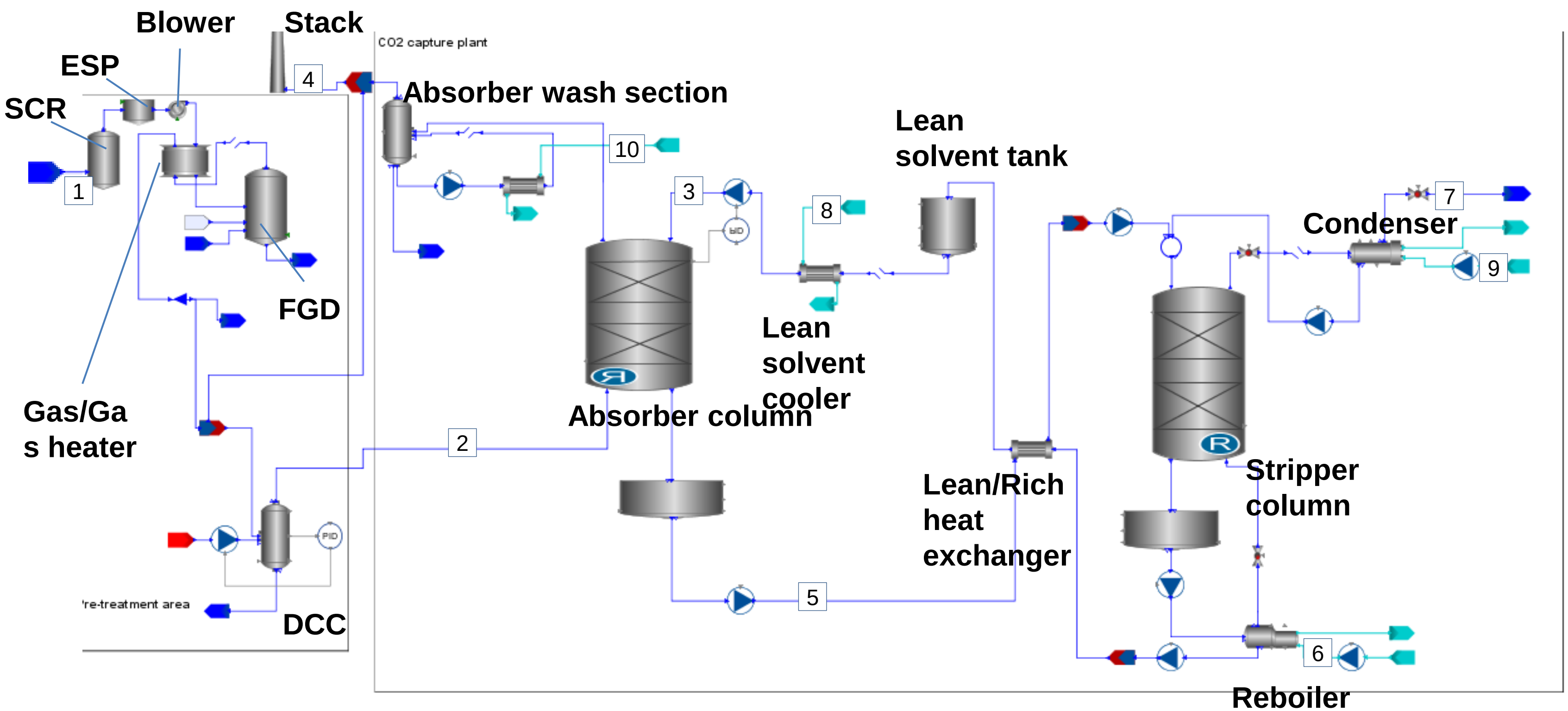
# Narrative

- Flue gas after the scrubber at Dunbar undergoes further pre-treatment. The pre-treatment area consists of a number of unit operations.
- A Selective Catalytic Reduction unit is used to capture 90% of NOx emissions.
- An Electrostatic precipitator is used to reduce virtually all the particulate emissions.
- A blower raises the pressure of the gas stream to overcome the pressure drops of the downstream systems (FGD, DCC  and absorber)
- A Flue gas desulphurisation (FGD) unit is required to reduce SOx emissions to acceptable levels. Some carbon dioxide is generated in the process and affects the mass balance.

# Narrative (continued)

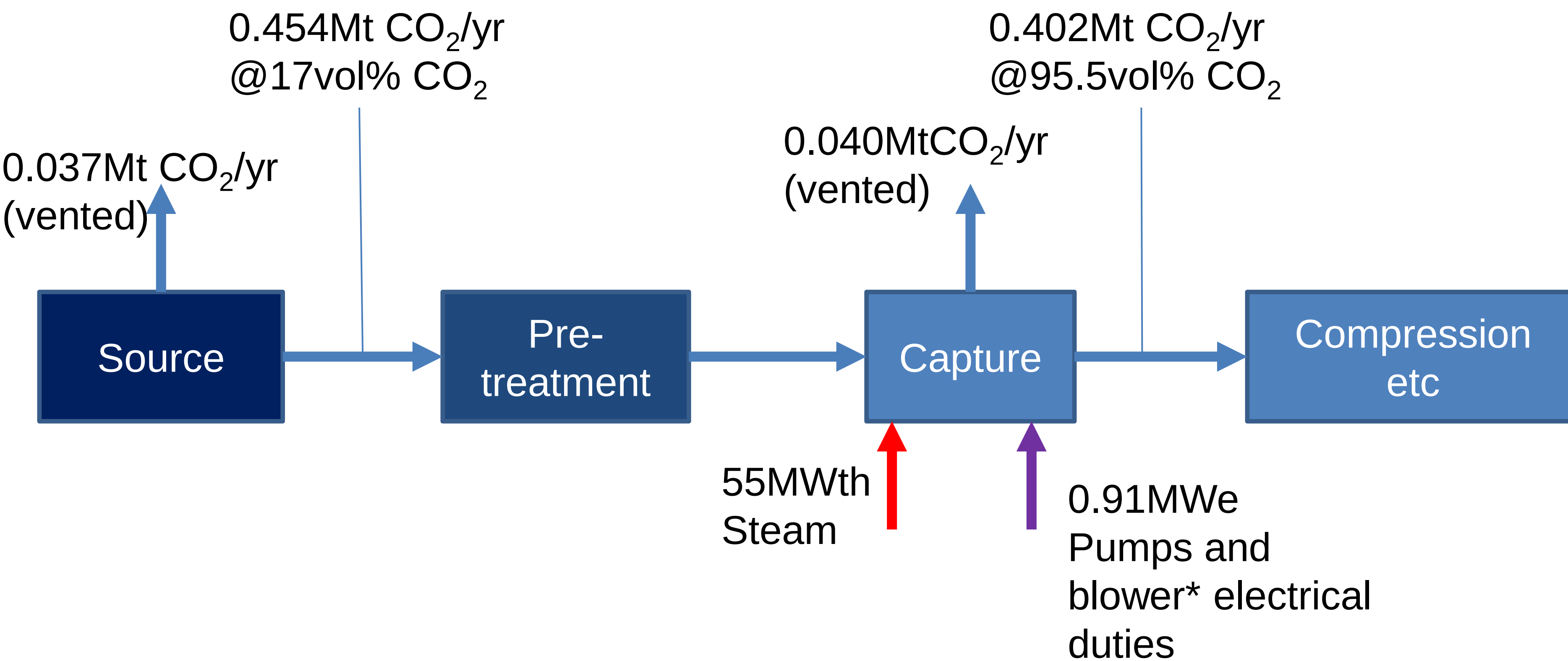
- The flue gas is  cooled in a direct contact cooler (DCC) to about 40°C. The gas is saturated with water vapour at those conditions. Some $CO_2$ is also dissolved in the quench water in the DCC. The overall balance results in the difference in the mass flows of $CO_2$ in Streams 1 and 2.
- The gas flows into the absorber where it is countercurrently contacted with MEA solvent. MEA chemically absorbs $CO_2$ and the resultant (rich) solvent is pumped from the absorber sumps through a lean/rich heat exchanger to the stripper column for regeneration. The duty required in the reboiler of the column for solvent regeneration is supplied by low-pressure steam (about 3.5bara). In the partial condenser of the stripper column, $CO_2$ is separated from water vapour before it is compressed in the downstream compressor units.
- The hot regenerated (lean) solvent heats up the cooler rich solvent in the lean/rich heat exchanger and is further cooled in a lean amine cooler before it flows to the absorber, completing the cycle.
- A buffer tank is used in the process where make-up solvent and/or water could be added to the process. The tank also provides additional flexibility to the process based on its capacity.
- Gas from the top of the absorber is cooled and vapourised amine solvent is captured in the absorber wash water section.

# Baseline Process Simulation of cement-MEA configuration




SCR – Selective Catalytic Reduction
ESP – Electrostatic Precipitator
FGD – Flue Gas Desulphurization
DCC – Direct Contact Cooler

# Baseline Process Simulation
# High Level Process/Energy flows




* Blower is assumed to provide 0.05bar pressure difference

# Stream Tables

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | CO2 SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS | LEAN AMINE | OUTLET GAS | RICH AMINE |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 19,398 | 9,633 | 732,204 | 7,120 | 711,569 |
| **MEA** | | 0 | 0.0 | 343487.0 | 0.0 | 343342.7 |
| **$CO_2$** | | 54,020 | 54,003 | 69,266 | 5,359 | 117,163 |
| **$N_2$** | | 113,949 | 146,347 | 0 | 114,923 | 23 |
| **$O_2$** | | 30,810 | 0 | 0 | 31,061 | 0 |
| **$SO_2$** | | 85 | 0 | 0 | 3 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 328 | 0 | 0 | 328 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 218,664 | 209,984 | 1,144,957 | 158,802 | 1,172,098 |
| **Temperature** | ºC | 55.5 | 40.9 | 40.8 | 40.0 | 59.3 |
| **Pressure** | bar(a) | 1.05 | 1.10 | 1.06 | 1.06 | 1.10 |

# Stream Tables (continued)

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 91,796 | 904 | 1,517,996 | 1,262,929 | 2,601,569 |
| **MEA** | | 0.0 | 0.4 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 47,884 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 23 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 91,796 | 48,812 | 1,517,996 | 1,262,929 | 2,601,569 |
| **Temperature** | $^oC$ | 127.5 | 39.8 | 9.9 | 25.0 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
# Baseline scenario - Process Conditions

| Description | Value |
| --- | --- |
| Number of trains of capture plant | 1 |
| Source vol % $CO_2$ | 16.7 |
| Site total $CO_2$ captureable (tonnes/year) | 491,000 |
| % site $CO_2$ captureable | 92.4 |
| Total reboiler heat duty (MWth) | 55.0 |
| Specific reboiler duty (GJ/ $tCO_2$) | 4.13 |
| Total electrical power requirement of capture plant pumps (MWe) | 0.58 |
| Electrical power requirement of blower* (MWe) | 0.33 |
| Total Cooling water required (tonnes/hr) | 5,382 |
| Total Capture plant site area required ($m^2$) | 7,850 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 307<br>SOx – 181 |
| After (ppm) | NOx – 40<br>SOx – 7 |

* Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Baseline scenario – Equipment list

| Summary | | | Equipment Sizing outputs | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 7.2 | Packing Height (m) | 21.2 | T/T Height (m) | 51.2 | 5,617,091 | 52 |
| Stripper | Diameter (m) | 5.6 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 2,155,438 | 20 |
| Reboiler | Heat Duty (MWth) | 55.0 | Steam flowrate (t/h) | 7.1 | | | 709,180 | 7 |
| Condenser | Cooling Duty (MWth) | 14.7 | Cooling water flowrate (t/h) | 97.4 | | | 317,889 | 3 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 49.9 | Heat transfer area per heat exchanger (m2) | 665.7 | Number of heat exchangers | 1 | 84,485 | 1 |
| Lean amine tank | Volume of tank (m3) | 5.0 | | | | | 295,803 | 3 |
| Lean amine cooler | Cooling Duty (MWth) | 35.3 | Heat transfer area per heat exchanger (m2) | 434.4 | Number of heat exchangers | 4 | 442,872 | 4 |
| Rich solvent pump | Total power requirement (kW) | 320.9 | Number of pumps required | 6 | | | 92,975 | 1 |
| Lean solvent pump | Total power requirement (kW) | 120.9 | Number of pumps required | 3 | | | 45,835 | 0 |
| Cooling water pumps | Total power requirement (kW) | 141.6 | Number of pumps required | 24 | | | 402,173 | 4 |
| Steam boiler | Capacity (t/h steam) | 91.8 | | | | | 659,950 | 6 |
| **Total equipment purchase cost (PCE)** | | | | | | | **10,823,690** | |

# Simulation Results
# Baseline scenario - Capital Expenditure (per train)

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 5,617,091 | 51.9 |
| Stripper | 2,155,438 | 19.9 |
| Reboiler | 709,180 | 6.6 |
| Condenser | 317,889 | 2.9 |
| Lean/Rich Heat Exchanger | 84,485 | 0.8 |
| Lean amine tank | 295,803 | 2.7 |
| Lean amine cooler | 442,872 | 4.1 |
| Rich solvent pump | 92,975 | 0.9 |
| Lean solvent pump | 45,835 | 0.4 |
| Cooling water pumps | 402,173 | 3.7 |
| Steam boiler | 659,950 | 6.1 |
| **Total equipment purchase cost (PCE)** | **10,823,690** | |

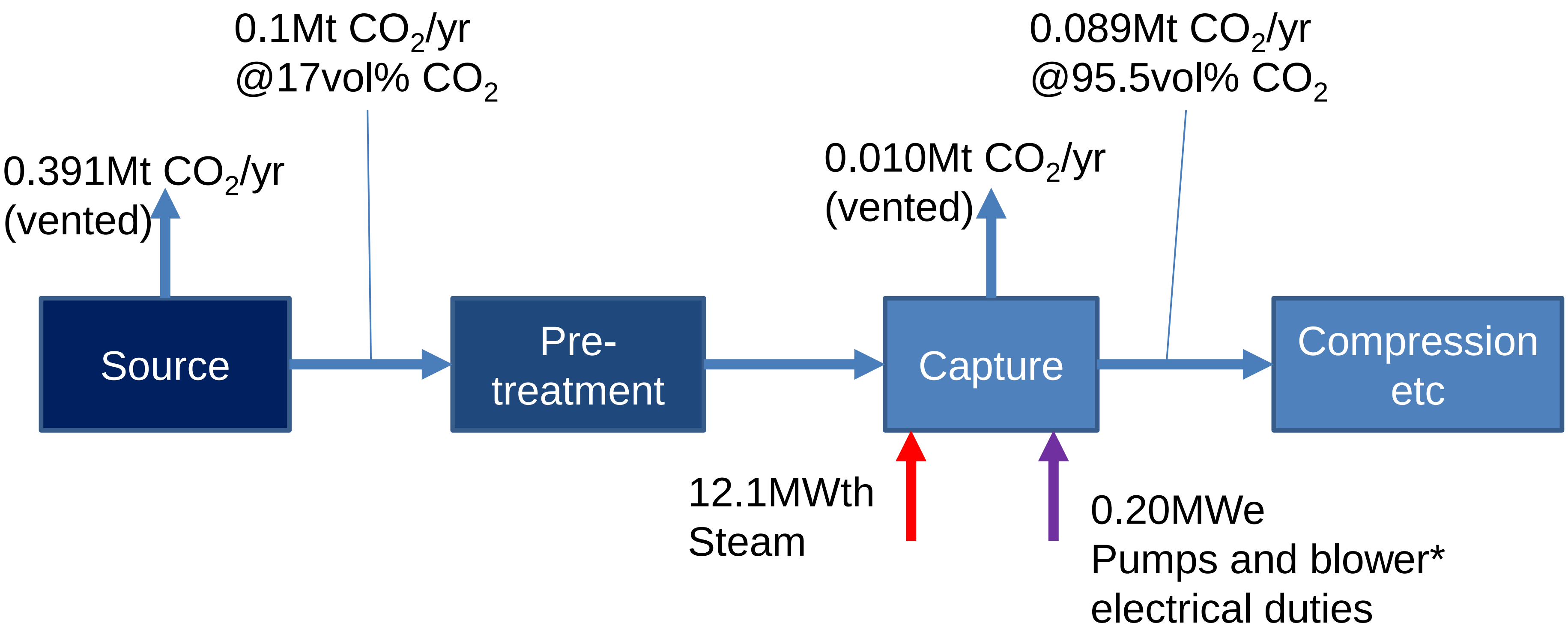
# Simulation Results
# Baseline scenario - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 10,823,690 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 34,635,808 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **68,682,808** |

# Simulation Results
# Baseline scenario - Operating Expenditure

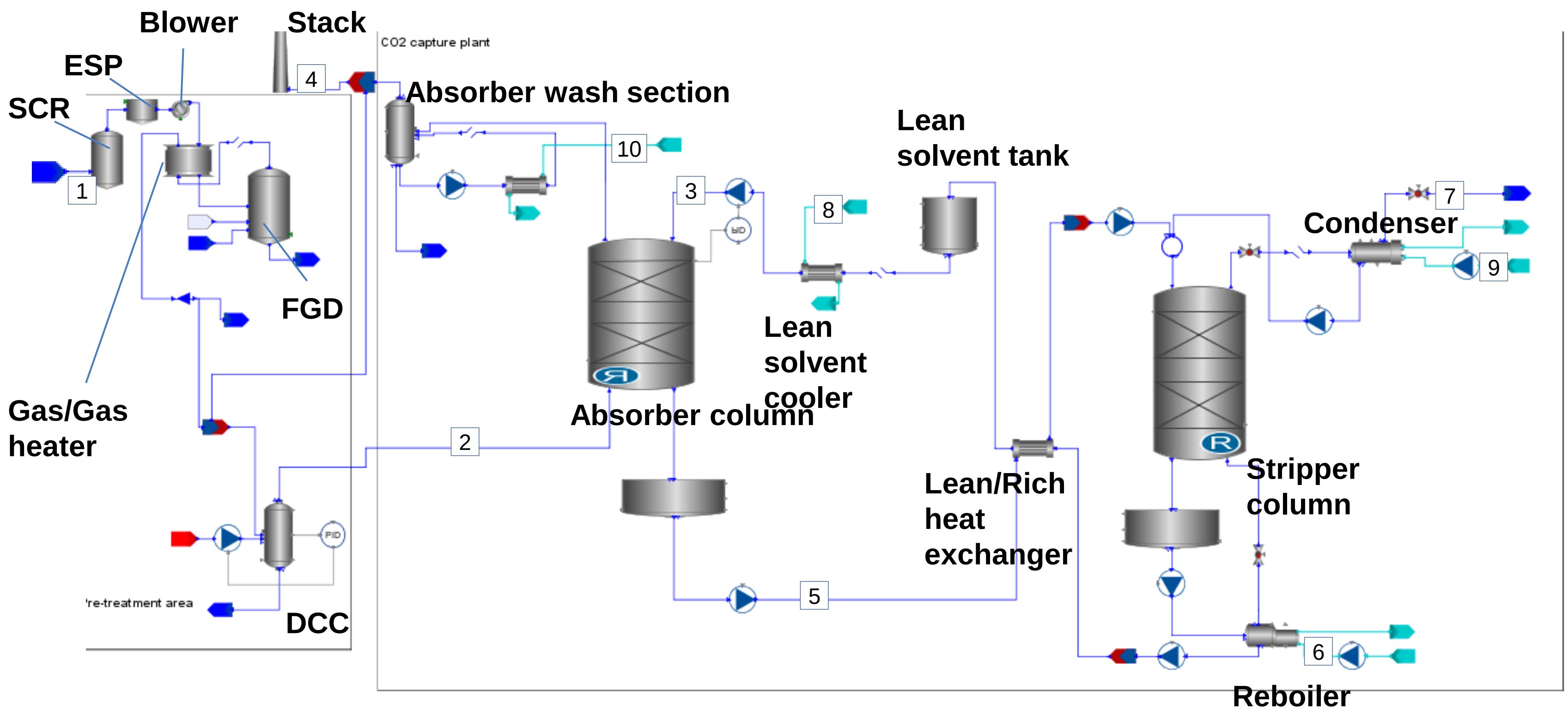
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 3,434,140 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 343,414 |
| Solvent make-up cost | | 2,029,691 |
| Pumps power cost | | 370,992 |
| Utilities - Steam costs | | 10,964,802 |
| Utilities - Cooling water costs | | 678,194 |
| **Total Variable costs** | | **14,387,093** |
| **OPEX** | | **17,821,233** |

# Scenario 2 Process Simulation
# High Level Process/Energy flows (Pilot plant scale)




* Blower is assumed to provide 0.05bar pressure difference

# Scenario 2 Process Simulation
# Pilot plant scale




**SCR – Selective Catalytic Reduction**
**ESP – Electrostatic Precipitator**
**FGD – Flue Gas Desulphurization**
**DCC – Direct Contact Cooler**

# Stream Tables
# Scenario 2

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | CO2 SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS | LEAN AMINE | OUTLET GAS | RICH AMINE |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 4,275 | 2,123 | 161,905 | 1,568 | 157,402 |
| **MEA** | | 0 | 0.0 | 75951.8 | 0.0 | 75920.3 |
| **$CO_2$** | | 11,905 | 11,901 | 15,316 | 1,181 | 25,873 |
| **$N_2$** | | 25,112 | 32,251 | 0 | 25,326 | 5 |
| **$O_2$** | | 6,790 | 0 | 0 | 6,845 | 0 |
| **$SO_2$** | | 19 | 0 | 0 | 1 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 72 | 0 | 0 | 72 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 48,188 | 46,275 | 253,173 | 34,995 | 259,201 |
| **Temperature** | ºC | 55.5 | 41.0 | 40.8 | 40.0 | 59.4 |
| **Pressure** | bar(a) | 1.05 | 1.10 | 1.07 | 1.07 | 1.10 |

# Stream Tables (continued)
# Scenario 2

| From: | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
|---|---|---|---|---|---|---|
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 20,269 | 199 | 336,367 | 279,261 | 568,555 |
| **MEA** | | 0.0 | 0.1 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 10,554 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 5 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 20,269 | 10,759 | 336,367 | 279,261 | 568,555 |
| **Temperature** | ºC | 127.5 | 39.8 | 9.9 | 25.0 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
# Scenario 2 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 1 |
| Source vol % $CO_2$ | 16.7 |
| Site total $CO_2$ captureable (tonnes/year) | 491,000 |
| % site $CO_2$ captureable | 20.4 |
| Total reboiler heat duty (MWth) | 12.1 |
| Specific reboiler duty (GJ/ $tCO_2$) | 4.13 |
| Total electrical power requirement of capture plant pumps (MWe) | 0.13 |
| Electrical power requirement of blower* (MWe) | 0.07 |
| Total Cooling water required (tonnes/hr) | 1,184 |
| Total Capture plant site area required ($m^2$) | 3,683 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 307<br>SOx – 181 |
| After (ppm) | NOx – 40<br>SOx – 7 |

* Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 2 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 3.4 | Packing Height (m) | 20.6 | T/T Height (m) | 50.6 | 1,474,564 | 48 |
| Stripper | Diameter (m) | 2.6 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 652,098 | 21 |
| Reboiler | Heat Duty (MWth) | 12.1 | Steam flowrate (t/h) | 1.6 | | | 154,903 | 5 |
| Condenser | Cooling Duty (MWth) | 3.2 | Cooling water flowrate (t/h) | 21.5 | | | 115,569 | 4 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 11.0 | Heat transfer area per heat exchanger (m2) | 146.8 | Number of heat exchangers | 1 | 21,101 | 1 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 134,924 | 4 |
| Lean amine cooler | Cooling Duty (MWth) | 7.8 | Heat transfer area per heat exchanger (m2) | 96.2 | Number of heat exchangers | 4 | 160,671 | 5 |
| Rich solvent pump | Total power requirement (kW) | 71.0 | Number of pumps required | 2 | | | 24,551 | 1 |
| Lean solvent pump | Total power requirement (kW) | 26.7 | Number of pumps required | 1 | | | 12,125 | 0 |
| Cooling water pumps | Total power requirement (kW) | 31.1 | Number of pumps required | 6 | | | 92,922 | 3 |
| Steam boiler | Capacity (t/h steam) | 20.3 | | | | | 206,920 | 7 |
| Total equipment purchase cost (PCE) | | | | | | | **3,050,347** | |

# Simulation Results
# Scenario 2 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 1,474,564 | 48 |
| Stripper | 652,098 | 21 |
| Reboiler | 154,903 | 5 |
| Condenser | 115,569 | 4 |
| Lean/Rich Heat Exchanger | 21,101 | 1 |
| Lean amine tank | 134,924 | 4 |
| Lean amine cooler | 160,671 | 5 |
| Rich solvent pump | 24,551 | 1 |
| Lean solvent pump | 12,125 | 0 |
| Cooling water pumps | 92,922 | 3 |
| Steam boiler | 206,920 | 7 |
| **Total equipment purchase cost (PCE)** | **3,050,347** | |

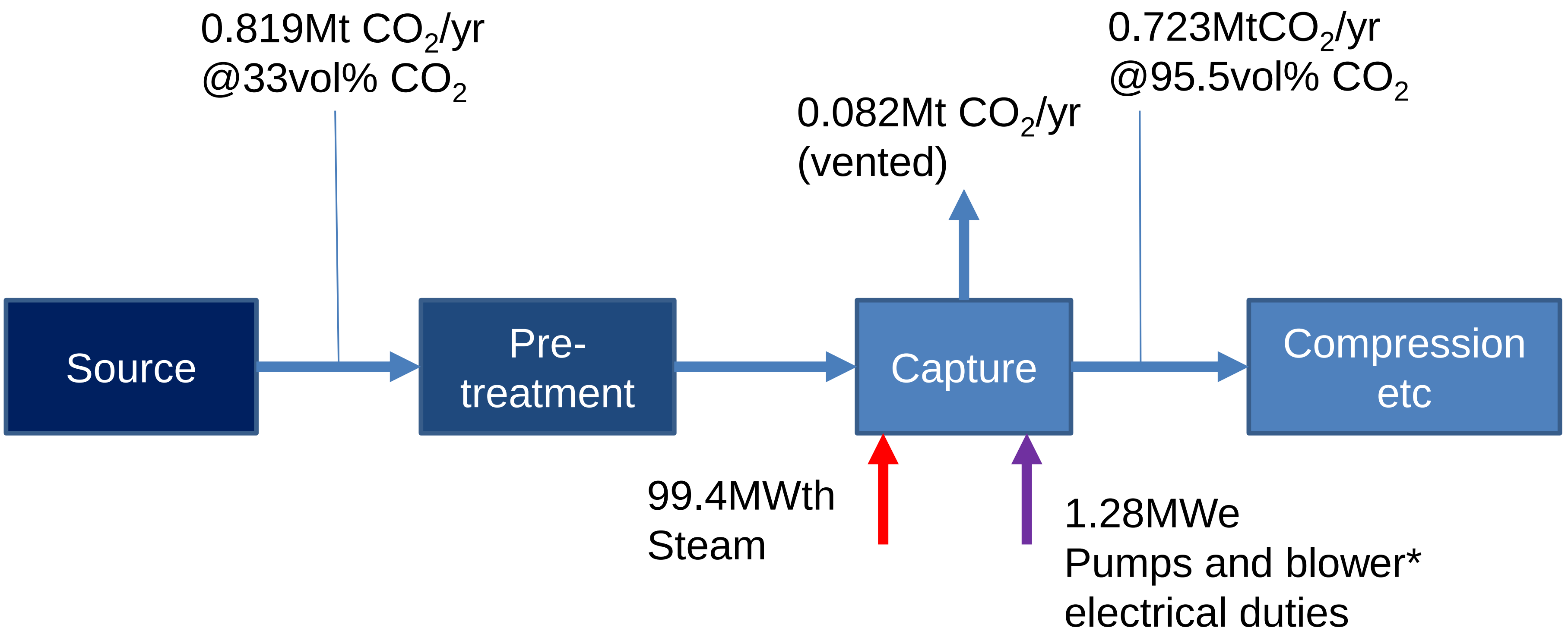
# Simulation Results
# Scenario 2 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 3,050,347 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 9,761,112 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **19,356,285** |

# Simulation Results
# Scenario 2 - Operating Expenditure

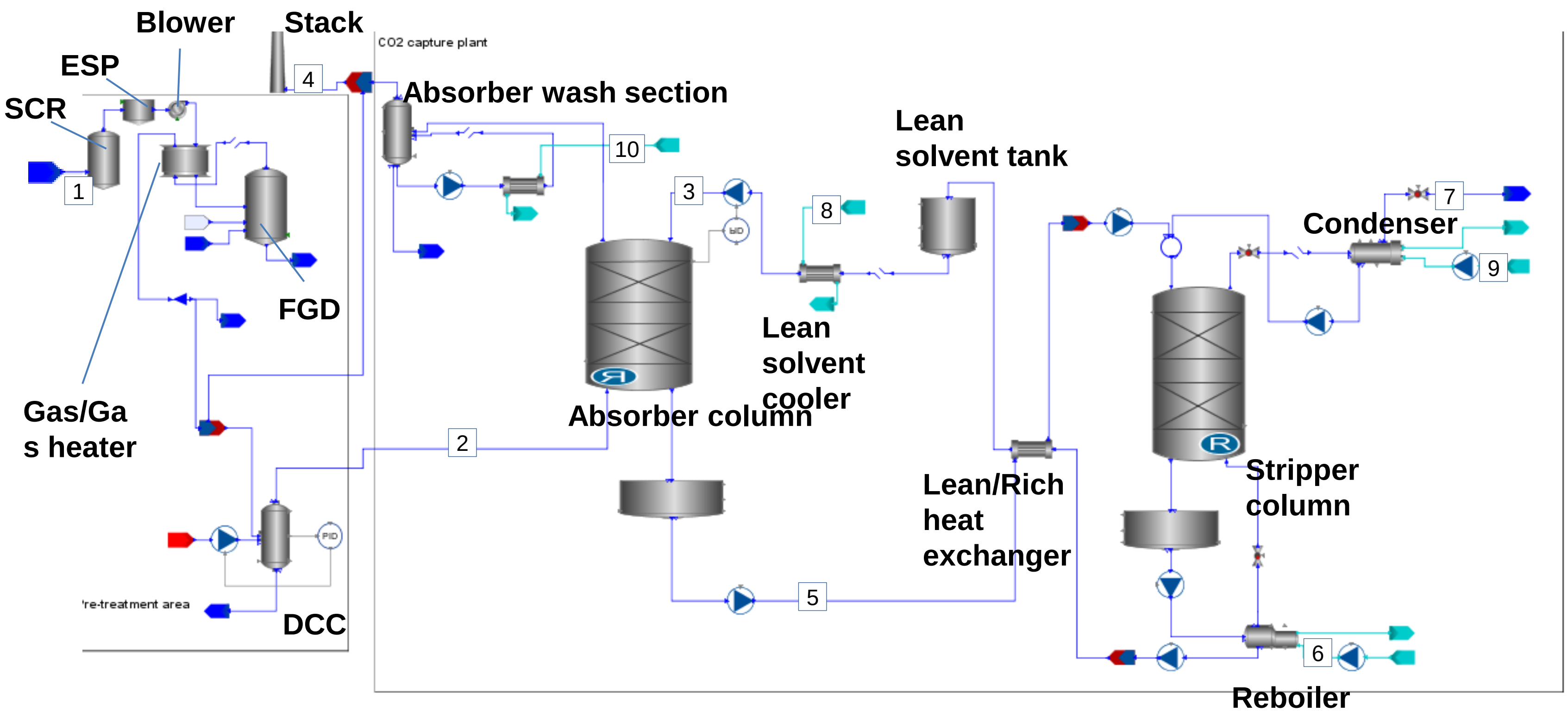
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 967,814 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 96,781 |
| Solvent make-up cost | | 447,369 |
| Pumps power cost | | 79,214 |
| Utilities - Steam costs | | 2,407,513 |
| Utilities - Cooling water costs | | 149,207 |
| **Total Variable costs** | | **3,180,083** |
| **OPEX** | | **4,147,898** |

## Scenario 3 Process Simulation
## High Level Process/Energy flows (33% CO2)




* Blower is assumed to provide 0.05bar pressure difference

# Scenario 3 Process Simulation
# 33% CO2 in Flue gas stream




**SCR – Selective Catalytic Reduction**
**ESP – Electrostatic Precipitator**
**FGD – Flue Gas Desulphurization**
**DCC – Direct Contact Cooler**

# Stream Tables
# Scenario 3

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | CO2 SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS | LEAN AMINE | OUTLET GAS | RICH AMINE |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 14,262 | 8,557 | 1,319,154 | 5,394 | 1,308,981 |
| **MEA** | | 0 | 0.0 | 618848.3 | 0.0 | 618749.0 |
| **$CO_2$** | | 97,589 | 97,539 | 124,826 | 9,720 | 210,866 |
| **$N_2$** | | 83,776 | 107,824 | 0 | 84,660 | 30 |
| **$O_2$** | | 22,652 | 0 | 0 | 22,882 | 0 |
| **$SO_2$** | | 78 | 0 | 0 | 2 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 241 | 0 | 0 | 241 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 218,664 | 213,919 | 2,062,828 | 122,906 | 2,138,626 |
| **Temperature** | ºC | 55.5 | 40.0 | 40.8 | 40.0 | 68.6 |
| **Pressure** | bar(a) | 1.05 | 1.10 | 1.07 | 1.07 | 1.10 |

# Stream Tables (continued)
# Scenario 3

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 165,806 | 1,624 | 3,428,791 | 2,291,706 | 1,451,653 |
| **MEA** | | 0.0 | 0.8 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 86,017 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 30 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 165,806 | 87,671 | 3,428,791 | 2,291,706 | 1,451,653 |
| **Temperature** | ºC | 127.5 | 39.8 | 9.9 | 25.0 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
# Scenario 3 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 1 |
| Source vol % $CO_2$ | 33 |
| Site total $CO_2$ captureable (tonnes/year) | 491,000 |
| % site $CO_2$ captureable | 167 |
| Total reboiler heat duty (MWth) | 99.4 |
| Specific reboiler duty (GJ/ tCO2) | 4.16 |
| Total electrical power requirement of capture plant pumps (MWe) | 0.98 |
| Electrical power requirement of blower* (MWe) | 0.30 |
| Total Cooling water required (tonnes/hr) | 7,172 |
| Total Capture plant site area required ($m^2$) | 8,490 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 307<br>SOx – 181 |
| After (ppm) | NOx – 48<br>SOx – 8 |

* Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 3 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 7.8 | Packing Height (m) | 16.9 | T/T Height (m) | 46.9 | 5,439,513 | 35 |
| Stripper | Diameter (m) | 7.6 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 3,594,429 | 23 |
| Reboiler | Heat Duty (MWth) | 99.4 | Steam flowrate (t/h) | 12.8 | | | 1,327,256 | 9 |
| Condenser | Cooling Duty (MWth) | 26.6 | Cooling water flowrate (t/h) | 176.8 | | | 570,122 | 4 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 72.0 | Heat transfer area per heat exchanger (m2) | 987.0 | Number of heat exchangers | 10 | 1,222,172 | 8 |
| Lean amine tank | Volume of tank (m3) | 180.3 | | | | | 406,640 | 3 |
| Lean amine cooler | Cooling Duty (MWth) | 79.7 | Heat transfer area per heat exchanger (m2) | 907.6 | Number of heat exchangers | 4 | 922,461 | 6 |
| Rich solvent pump | Total power requirement (kW) | 590.3 | Number of pumps required | 10 | | | 163,812 | 1 |
| Lean solvent pump | Total power requirement (kW) | 220.8 | Number of pumps required | 5 | | | 80,098 | 1 |
| Cooling water pumps | Total power requirement (kW) | 170.0 | Number of pumps required | 32 | | | 536,020 | 3 |
| Steam boiler | Capacity (t/h steam) | 165.8 | | | | | 1,128,714 | 7 |
| Total equipment purchase cost (PCE) | | | | | | | **15,391,239** | |

# Simulation Results
# Scenario 3 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 5,439,513 | 35.3 |
| Stripper | 3,594,429 | 23.4 |
| Reboiler | 1,327,256 | 8.6 |
| Condenser | 570,122 | 3.7 |
| Lean/Rich Heat Exchanger | 1,222,172 | 7.9 |
| Lean amine tank | 406,640 | 2.6 |
| Lean amine cooler | 922,461 | 6.0 |
| Rich solvent pump | 163,812 | 1.1 |
| Lean solvent pump | 80,098 | 0.5 |
| Cooling water pumps | 536,020 | 3.5 |
| Steam boiler | 1,128,714 | 7.3 |
| **Total equipment purchase cost (PCE)** | **15,391,239** | |

# Simulation Results
## Scenario 3 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 15,391,239 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 49,251,963 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **97,666,643** |

# Simulation Results
# Scenario 3 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 4,883,332 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 488,333 |
| Solvent make-up cost | | 3,645,530 |
| Pumps power cost | | 623,859 |
| Utilities - Steam costs | | 19,805,202 |
| Utilities - Cooling water costs | | 903,691 |
| **Total Variable costs** | | **25,466,615** |
| **OPEX** | | **30,349,947** |

# Comparison of costs between scenarios

| | Baseline Scenario #1 | Sensitivity "Pilot/Demo" (Scenario #2) | Sensitivity "High CO2 concentration" (Scenario #3) |
|---|---|---|---|
| Source $CO_2$ | 0.45 Mt/yr | 0.10 Mt/yr | 0.82Mt/yr |
| Equipment cost | £11m | £3m | £15m |
| Total fixed cost | £58m | £16m | £82m |
| Annual opex (incl. energy) | £18m/yr | £4m/yr | £30m/yr |
| Reboiler Heat Duty $MW_{th}$ | 55.0 | 12.1 | 99.4 |
| Power /$MW_e$ | 0.91 | .020 | 1.28 |

# Insights from process simulation

- Capture with a 1st generation amine solvent at a UK cement plant is feasible.
- The overall capital cost is *ca*. 6 times the cost of the main pieces of equipment.
- The largest of the equipment cost items modelled is the absorber.
- The largest operating cost modelled is for steam.
- Therefore capture technology development should focus on reducing the costs of absorber, and/or the amount of steam required, and simplifying retrofit installation.
- Water availability should be discussed with stakeholders.

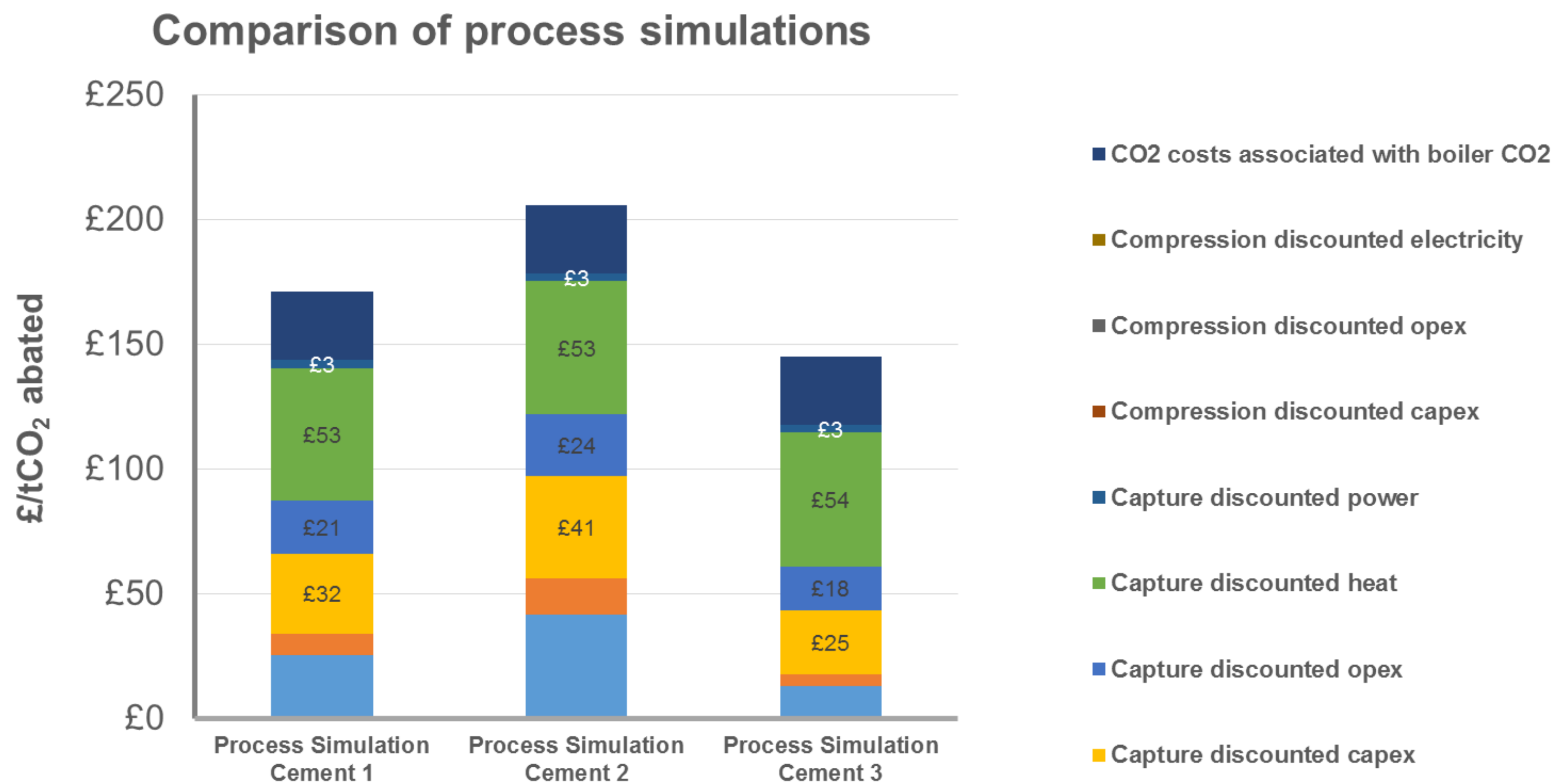
# Summary outputs from process simulations and techno-economic modelling can be compared and used to prioritise model refinements.

| Parameter | Techno-economics "Baseline" | Process simulation "Baseline" |
|---|---|---|
| Input flue gas $MtCO_2$/yr | 0.49 Mt/yr | 0.45 Mt/yr |
| Abated $MtCO_2$/yr | 0.33 Mt/yr | Not calculated directly |
| Capex | Capture only: £109m (£138m incl. pre-treatment) | Capture only: £69m |
| Non-energy opex | £7m/yr | £6m/yr |
| Heat | 47 MW | 55 MW |
| Power | 1.32 MW | 0.91 MW |

# Issues emerging

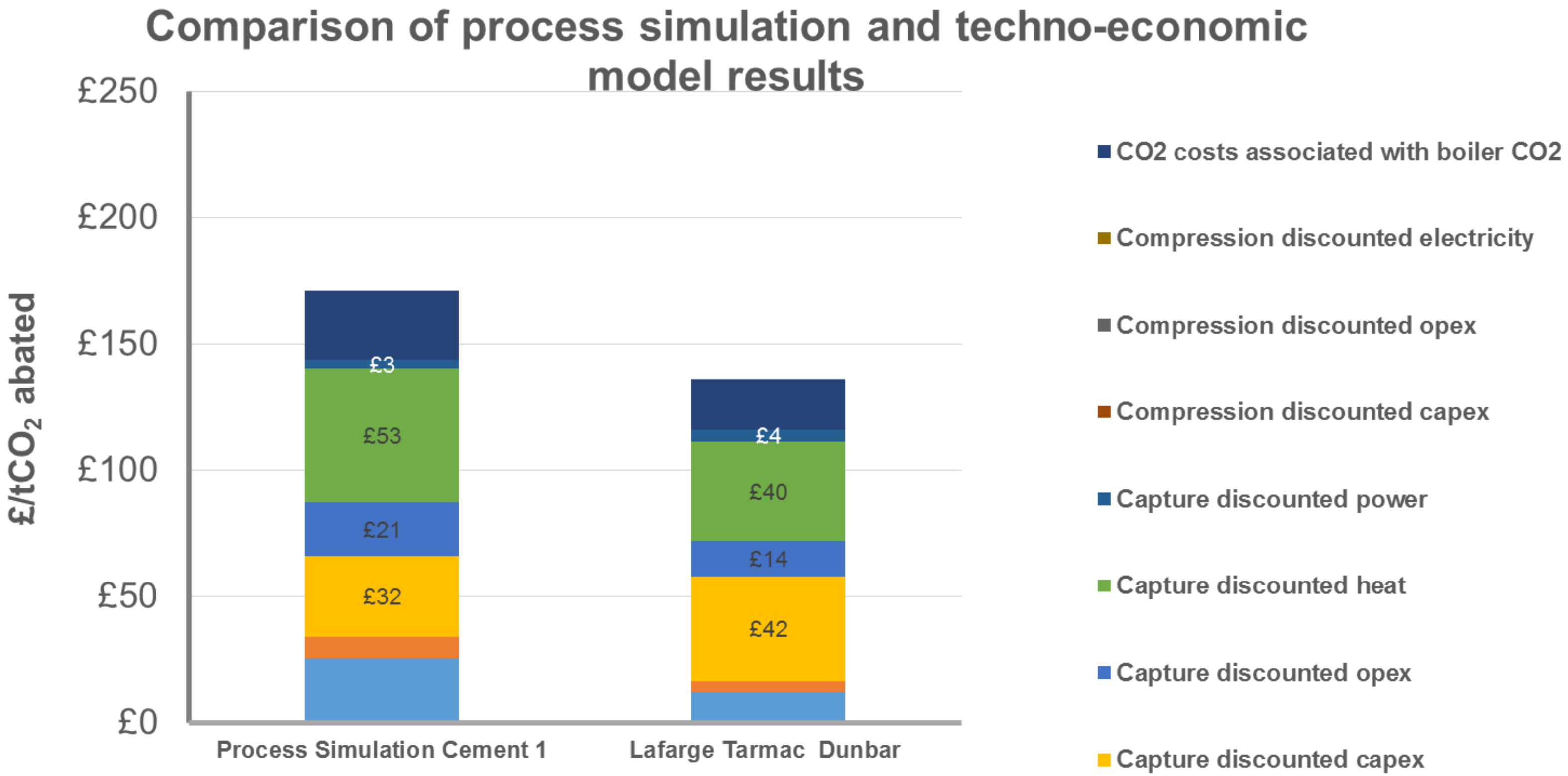
- It is not surprising that opposite approaches to estimating costs of capture do not agree completely.
- If a policy objective is to implement close to full scale capture (i.e. 0.5 Mt/yr) at a UK cement works by 2025, then the key barrier to overcome is experience of capture of cement $CO_2$ streams at an appropriate scale.
- A plausible development strategy to overcome this barrier could be to begin with a pilot project capturing ca. 0.1 $MtCO_2$/yr by 2020 using a single train 1$^{st}$ generation amine system, although other technologies are feasible.
- Following successful experience with this, it should be possible to build additional capture capacity by 2025 to capture 0.5 $MtCO_2$/yr or reduce unit costs by employing second generation chemical solvents or solid looping.

# Comparison of abatement costs for cement-MEA process simulation scenarios.




| Process Simulation Cement 1 | Process Simulation Cement 2 | Process Simulation Cement 3 |
|---|---|---|
| 17% $CO_2$ 0.45 Mt/yr input 0.31 Mt/yr abated | 17% $CO_2$ 0.1 Mt/yr input 0.07 Mt/yr abated | 33% $CO_2$ 0.82 Mt/yr input 0.56 Mt/yr abated |

# Comparison of cost breakdown for cement 2025 demonstration from process simulation and techno-economic modelling




Process simulation
17% $CO_2$
0.45 Mt/yr input
0.3 Mt/yr abated

Techno-economics
24% $CO_2$
0.45 Mt/yr input
0.35 Mt/yr abated

# Four case studies on process simulation

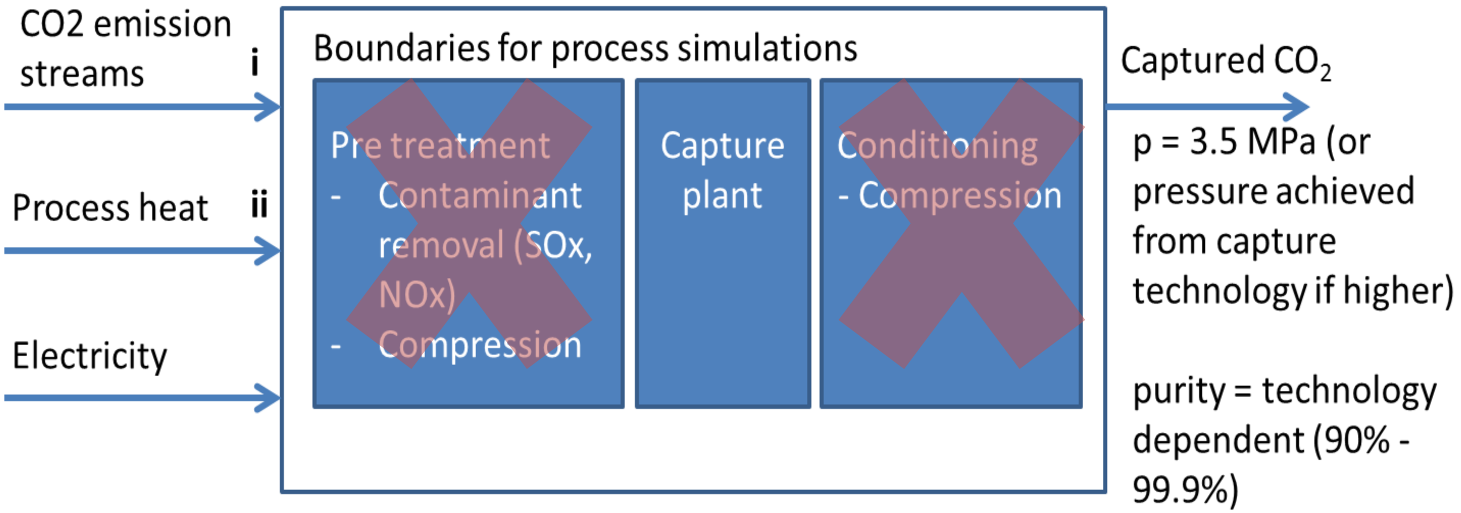
1. Grangemouth oil refinery with MEA capture
2. Lafarge Tarmac Dunbar cement works with MEA capture
3. Tata Scunthorpe iron and steel plant with MEA capture at CHP unit
4. GrowHow ammonia production site with $CO_2$ compression

# Scenarios for Iron and Steel MEA configuration

- Several technology options were considered for the process simulation based on literature review, techno-economic modelling and discussions with stakeholders.
- Based on data availability within the constraints of the study, stakeholders agreed to the process simulation should focus on retrofitting MEA-based post-combustion capture to a CHP plant burning (primarily) Blast Furnace Gas, corresponding to the output of one of the four Blast Furnaces at the Tata Scunthorpe site.
- It was assumed that this site would meet emissions controls on SOx and NOx.
- Priorities for uncertainty analysis included the impact of scale and concentration of $CO_2$ in the flue gas on capture design. Therefore three scenarios were developed for process simulation. These correspond to baseline, low, and high $CO_2$ concentrations, *assuming constant overall flue gas input to the capture facility*.
- Flue gas assumptions were reviewed with Tata Steel.

| Scenario | $CO_2$ concentration | Mt/yr |
|---|---|---|
| 1- Baseline | 20% | 2.53 |
| 2- Low | 15% | 1.95 |
| 3- High | 25% | 3.07 |

# Scope of study




**i** - only process emissions (excluding power and heat production emissions)
- in case multiple stack emissions are combined, piping etc to bring together is included

**ii** - including integration cost in case of process heat recovery
- assume natural gas boiler if no process heat is available (default assumption)

Further Notes

- Compression plant not simulated
- Pre-treatment processes simulated but not costed

# Detailed assumptions

- Requirements to convey flue sources to capture plant are ignored
- A basic MEA $CO_2$ capture plant configuration is selected for the study (complex configurations such as split-flow considerations are ignored)
- An optimal lean loading was estimated to take into account differences in inlet $CO_2$ concentration. This was based on a 1500m$^2$ lean/rich amine heat exchanger area and overall heat transfer coefficient of 6000W/m$^2$K
- The absorber packing height was sized to achieve a rich loading of 0.488 mol $CO_2$/mol MEA. The rich loading value used was assumed to be representative of the physical constraint on the capacity of the MEA solvent in chemical absorption processes.
- The column models were rate-based, distributed models. A trade-off between simulation performance (times) and accuracy was made reducing the number of discretization elements. This resulted in slight mass imbalances for certain components of up to about 1%.
- $O_2$/$N_2$ composition in flue gas streams are lumped as inert material
- $SO_X$ and $NO_X$ compositions are not considered at capture plant boundary
- Solvent degradation effects ignored
- As degradation rates are not estimated, no reclaimer is modelled or costed in this unit.
- All costs provided are overall costs except those provided in the equipment list (costs per train).

# Detailed assumptions (continued)

- The site area required for the capture plant is based on published work[1] and scaled as linearly dependent on the absorber diameter. This was done based on the fact that the absorber typically presents the largest footprint on the plant.
- In the pre-treatment area, the blower was assumed to provide a pressure increase of 0.05bar to overcome pressure drops in the downstream units.
- $CO_2$ emissions from the standalone steam boiler are not considered in this study.
- Cooling water requirements for the absorber wash water section are rough estimates. A trade-off is required between how much (and under what conditions) wash water is supplied to the wash section and the required cooling water flowrate. The optimisation of this scheme was outside the scope of this study and as a result was not carried out. A fixed flowrate of wash water was assumed and the cooling water flowrate to cool that stream is estimated.
- It should be noted that the same “hand factors” were used to estimate the total fixed capital costs from the total equipment purchase cost for all scales of the capture plants modelled. It may be more appropriate to employ different factors for different scales of plants.
- DECC’s 2025 fuel and electricity prices used for all the demo scenarios
- All other costs are based on Q3 2013. These could be translated to the same basis by assuming an appropriate discount rate.

[1] IEAGHG (2012). $CO_2$ capture at gas-fired power plants, Report No: 2012/8.

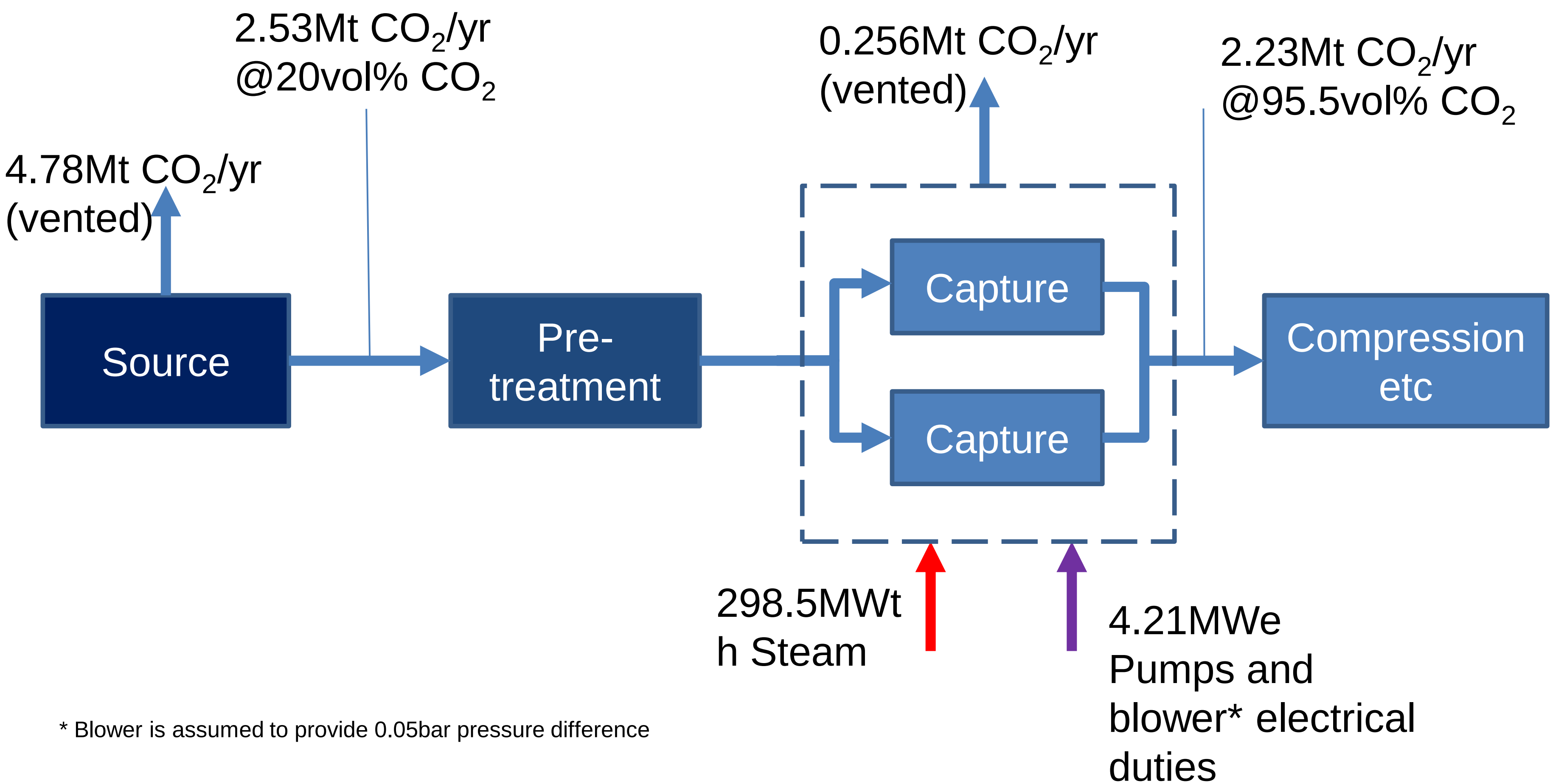
# Narrative

- Flue gas from combusted blast furnace gas first undergoes pre-treatment. The pre-treatment area consists of a number of unit operations.
- A Selective Catalytic Reduction unit is used to capture 90% of NOx emissions.
- An Electrostatic precipitator is used to reduce virtually all the particulate emissions.
- A blower raises the pressure of the gas stream to overcome the pressure drops of the downstream systems (FGD, DCC  and absorber)
- A Flue gas desulphurisation (FGD) unit is required to reduce SOx emissions to acceptable levels. Some carbon dioxide is generated in the process and affects the mass balance. This is seen in the difference in the mass flows of $CO_2$ in Streams 1 and 2.

# Narrative (continued)

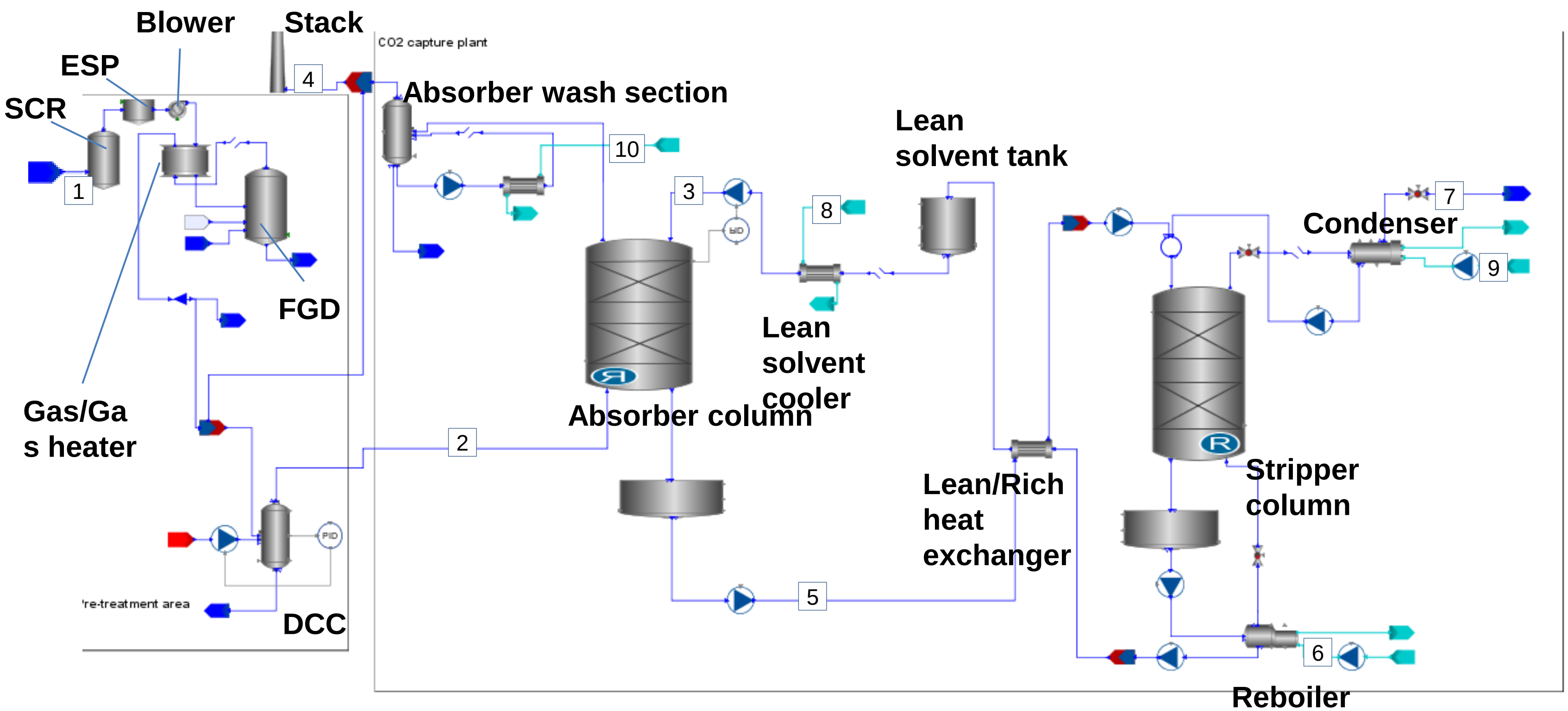
- The flue gas is  cooled in a direct contact cooler (DCC) to about 40°C. The gas is saturated with water vapour at those conditions. Some $CO_2$ is also dissolved in the quench water in the DCC. The overall balance results in the difference in the mass flows of $CO_2$ in Streams 1 and 2.
- The gas flows into the absorber where it is countercurrently contacted with MEA solvent. MEA chemically absorbs $CO_2$ and the resultant (rich) solvent is pumped from the absorber sumps through a lean/rich heat exchanger to the stripper column for regeneration. The duty required in the reboiler of the column for solvent regeneration is supplied by low-pressure steam (about 3.5bara). In the partial condenser of the stripper column, $CO_2$ is separated from water vapour before it is compressed in the downstream compressor units.
- The hot regenerated (lean) solvent heats up the cooler rich solvent in the lean/rich heat exchanger and is further cooled in a lean amine cooler before it flows to the absorber, completing the cycle.
- A buffer tank is used in the process where make-up solvent and/or water could be added to the process. The tank also provides additional flexibility to the process based on its capacity.
- Gas from the top of the absorber is cooled and vapourised amine solvent and other solvent derivatives are captured in the absorber wash water section.

# Baseline Process Simulation
# High Level Process/Energy flows




* Blower is assumed to provide 0.05bar pressure difference

# Baseline Process Simulation




SCR – Selective Catalytic Reduction
ESP – Electrostatic Precipitator
FGD – Flue Gas Desulphurization
DCC – Direct Contact Cooler

# Inputs - Baseline

- Two $CO_2$ capture trains
- 30 wt% MEA solvent
- 90% capture target
- >95% purity $CO_2$ by volume
- Absorber operating at atmospheric pressure
- Mellapak 250Y structured packing used in Absorber and stripper columns
- Heat exchanger overall heat transfer coefficient – 6000W/$m^2$K
- Heat exchanger area – 1500$m^2$
- Maximum flow in heat exchanger – 2500$m^3$/hr
- Lean loading specification – 0.233 mol $CO_2$/mol MEA
- Stripper operating pressure ~ 1.67bara
- Steam pressure 3.5bara
- Cooling water temperature is assumed to be 10°C
- 8400 hours of continuous operation assumed in a year

# Stream Tables
# Baseline scenario

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | CO2 SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS* | LEAN AMINE* | OUTLET GAS | RICH AMINE* |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 52,590 | 23,479 | 1,569,113 | 54,047 | 1,483,211 |
| **MEA** | | 0 | 0.0 | 724679.2 | 0.0 | 724054.5 |
| **$CO_2$** | | 300,763 | 150,419 | 121,805 | 30,418 | 254,639 |
| **$N_2$** | | 660,244 | 344,617 | 0 | 675,478 | 49 |
| **$O_2$** | | 26,882 | 0 | 0 | 13,639 | 0 |
| **$SO_2$** | | 288 | 0 | 0 | 4 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 1,040,904 | 518,515 | 2,415,597 | 773,600 | 2,461,953 |
| **Temperature** | ºC | 60.0 | 41.0 | 40.8 | 48.4 | 56.4 |
| **Pressure** | bar(a) | 1.01 | 1.10 | 1.07 | 1.07 | 1.10 |

* This value corresponds to a single train

# Stream Tables (continued)
# Baseline scenario

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 498,087 | 5,015 | 6,292,753 | 6,747,539 | 18,113,245 |
| **MEA** | | 0.0 | 2.5 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 265,607 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 98 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 498,087 | 270,722 | 6,292,753 | 6,747,539 | 18,113,245 |
| **Temperature** | ºC | 127.5 | 39.8 | 9.9 | 25.0 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
## Baseline scenario - Process Conditions

| Description | Value |
| --- | --- |
| Number of trains of capture plant | 2 |
| Source vol % $CO_2$ | 20 |
| Site total $CO_2$ captureable (tonnes/year) | 7,305,903 |
| % site $CO_2$ captureable | 34.6 |
| Total reboiler heat duty (MWth) | 298.5 |
| Reboiler Specific duty (GJ/t $CO_2$) | 4.05 |
| Lean loading (mol $CO_2$/mol MEA) | 0.233 |
| Total electrical power requirement of capture plant pumps (MWe) | 2.68 |
| Electrical power requirement of blower* (MWe) | 1.53 |
| Total Cooling water required (tonnes/hr) | 62,307 |
| Total Capture plant site area required ($m^2$) | 24,500 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.07 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 100<br>SOx – 131 |
| After (ppm) | NOx – 6<br>SOx – 6 |

* Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Baseline scenario – Equipment list

| Summary | | Equipment Sizing outputs | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 11.2 | Packing Height (m) | 21.9 | T/T Height (m) | 51.9 | 14,239,396 | 52 |
| Stripper | Diameter (m) | 9.4 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 6,047,594 | 22 |
| Reboiler | Heat Duty (MWth) | 149.3 | Steam flowrate (t/h) | 19.2 | | | 2,366,638 | 9 |
| Condenser | Cooling Duty (MWth) | 39.2 | Cooling water flowrate (t/h) | 260.3 | | | 866,727 | 3 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 112.8 | Heat transfer area per heat exchanger (m2) | 1500.0 | Number of heat exchangers | 1 | 181,250 | 1 |
| Lean amine tank | Volume of tank (m3) | 405.8 | | | | | 443,206 | 2 |
| Lean amine cooler | Cooling Duty (MWth) | 73.1 | Heat transfer area per heat exchanger (m2) | 901.1 | Number of heat exchangers | 4 | 915,408 | 3 |
| Rich solvent pump | Total power requirement (kW) | 671.9 | Number of pumps required | 12 | | | 191,594 | 1 |
| Lean solvent pump | Total power requirement (kW) | 251.2 | Number of pumps required | 6 | | | 94,688 | 0 |
| Cooling water pumps | Total power requirement (kW) | 418.6 | Number of pumps required | 30 | | | 492,945 | 2 |
| Steam boiler | Capacity (t/h steam) | 249.0 | | | | | 1,655,921 | 6 |
| **Total equipment purchase cost (PCE)** | | | | | | | **27,495,368** | |

[1]T/T height – tan to tan height representing the height from the top to the bottom of the column vessel.

# Simulation Results
## Baseline scenario - Capital Expenditure (per train)

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 28,478,793 | 51.8 |
| Stripper | 12,095,188 | 22.0 |
| Reboiler | 4,733,277 | 8.6 |
| Condenser | 1,733,454 | 3.2 |
| Lean/Rich Heat Exchanger | 362,501 | 0.7 |
| Lean amine tank | 886,413 | 1.6 |
| Lean amine cooler | 1,830,815 | 3.3 |
| Rich solvent pump | 383,187 | 0.7 |
| Lean solvent pump | 189,375 | 0.3 |
| Cooling water pumps | 985,890 | 1.8 |
| Steam boiler | 3,311,843 | 6.0 |
| **Total equipment purchase cost (PCE)** | **54,990,735** | |

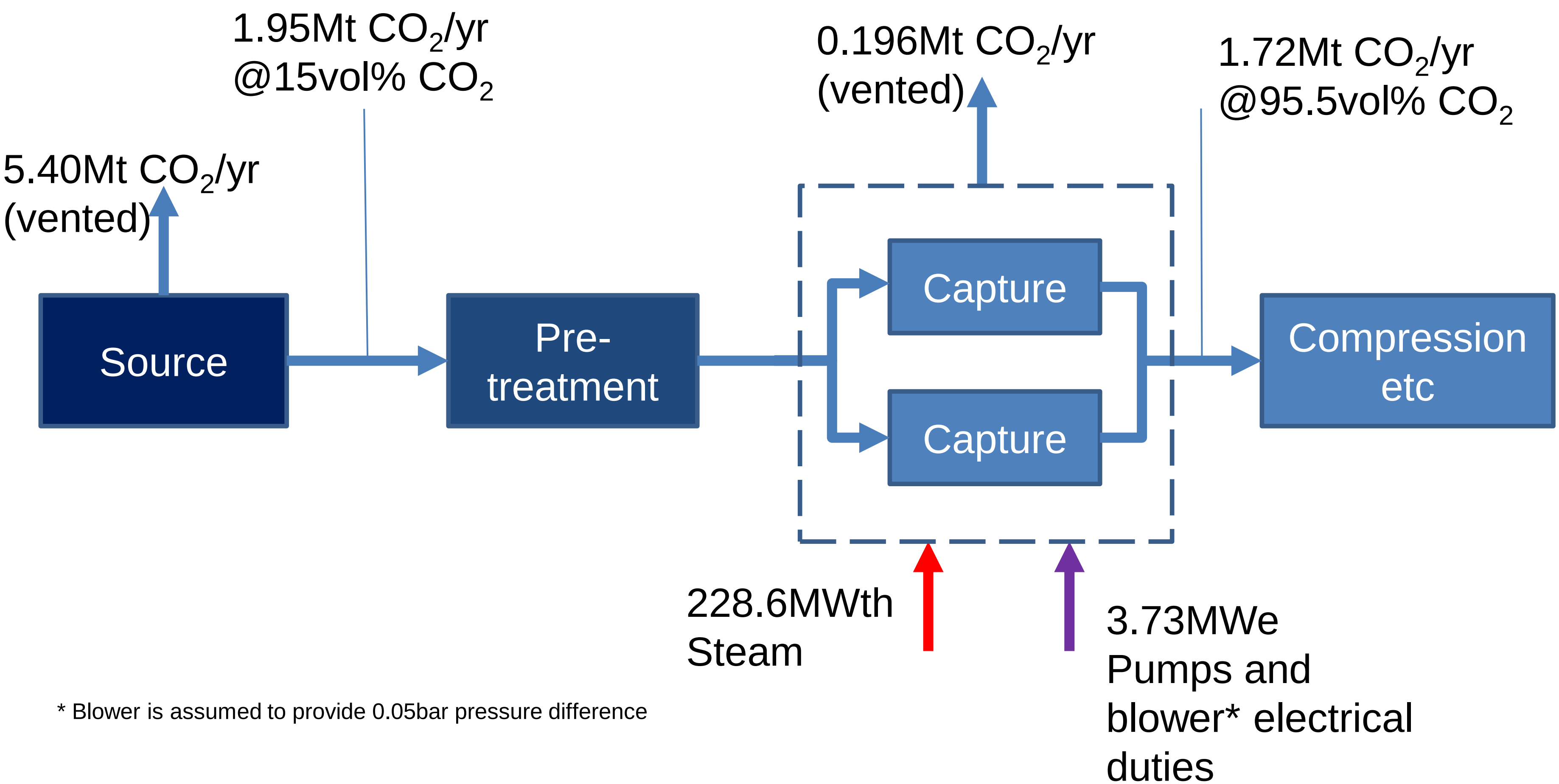
# Simulation Results
## Baseline scenario - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 54,990,735 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 175,970,353 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **348,949,210** |

# Simulation Results
## Baseline scenario - Operating Expenditure

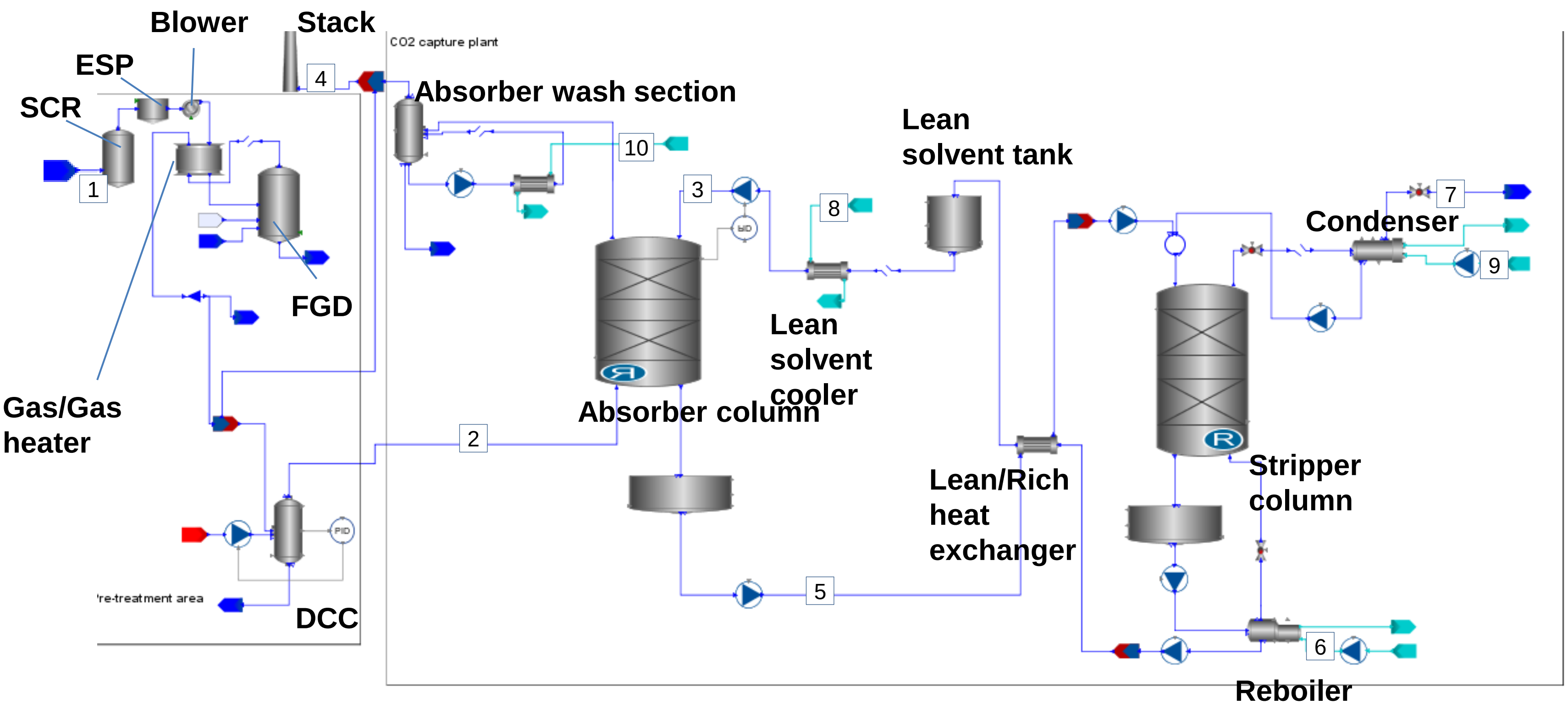
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 17,447,460 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 1,744,746 |
| Solvent make-up cost | | 11,257,106 |
| Pumps power cost | | 1,706,298 |
| Utilities - Steam costs | | 59,495,549 |
| Utilities - Cooling water costs | | 3,925,346 |
| **Total Variable costs** | | **78,129,044** |
| **OPEX** | | **95,576,505** |

# Scenario 2 Process Simulation
# High Level Process/Energy flows (15% $CO_2$)




* Blower is assumed to provide 0.05bar pressure difference

# Scenario 2 Process Simulation
# 15% $CO_2$ in combusted Blast Furnace gas stream




**SCR – Selective Catalytic Reduction**
**ESP – Electrostatic Precipitator**
**FGD – Flue Gas Desulphurization**
**DCC – Direct Contact Cooler**

# Inputs – Scenario 2

- Two $CO_2$ capture trains
- 30 wt% MEA solvent
- 90% capture target
- >95% purity $CO_2$ by volume
- Absorber operating at atmospheric pressure
- Mellapak 250Y structured packing used in Absorber and stripper columns
- Heat exchanger overall heat transfer coefficient – 6000W/m$^2$K
- Heat exchanger area – 1500m$^2$
- Maximum flow in heat exchanger – 2500m$^3$/hr
- Lean loading specification – 0.239 mol $CO_2$/mol MEA
- Stripper operating pressure ~ 1.67bara
- Steam pressure 3.5bara
- Cooling water temperature is assumed to be 10°C
- 8400 hours of continuous operation assumed in a year

# Stream Tables
# Scenario 2

| From: | | $CO_2$ SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
|---|---|---|---|---|---|---|
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS* | LEAN AMINE* | OUTLET GAS | RICH AMINE* |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 57,474 | 24,010 | 1,250,203 | 49,492 | 1,177,230 |
| **MEA** | | 0 | 0.0 | 578454.2 | 0.0 | 577950.5 |
| **$CO_2$** | | 232,022 | 116,061 | 99,523 | 23,310 | 202,180 |
| **$N_2$** | | 721,573 | 376,571 | 0 | 738,144 | 42 |
| **$O_2$** | | 29,380 | 0 | 0 | 14,894 | 0 |
| **$SO_2$** | | 315 | 0 | 0 | 5 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 1,040,904 | 516,642 | 1,928,181 | 825,858 | 1,957,403 |
| **Temperature** | ºC | 60.0 | 40.9 | 40.8 | 45.5 | 53.2 |
| **Pressure** | bar(a) | 1.01 | 1.10 | 1.07 | 1.07 | 1.10 |

* This value corresponds to a single train

# Stream Tables (continued)
# Scenario 2

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 381,343 | 3,871 | 4,248,205 | 5,540,060 | 16,093,935 |
| **MEA** | | 0.0 | 1.9 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 205,275 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 85 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 381,343 | 209,233 | 4,248,205 | 5,540,060 | 16,093,935 |
| **Temperature** | ºC | 127.5 | 39.8 | 9.9 | 25.0 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
# Scenario 2 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 2 |
| Source vol % $CO_2$ | 15 |
| Site total $CO_2$ captureable (tonnes/year) | 7,305,903 |
| % site $CO_2$ captureable | 26 |
| Total reboiler Heat duty (MWth) | 228.6 |
| Reboiler Specific duty (GJ/t CO2) | 4.01 |
| Lean loading (mol CO2/mol MEA) | 0.239 |
| Total electrical power requirement of capture plant pumps (MWe) | 2.16 |
| Electrical power requirement of blower* (MWe) | 1.57 |
| Cooling water required (tonnes/hr) | 25,882 |
| Capture plant site area required ($m^2$) | 24,460 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.08 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 100<br>SOx – 123 |
| After (ppm) | NOx – 6<br>SOx – 2 |

* Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 2 – Equipment list

| Summary | | Equipment Sizing outputs | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 11.2 | Packing Height (m) | 20.2 | T/T Height (m) | 50.2 | 13,367,503 | 54 |
| Stripper | Diameter (m) | 9.4 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 6,047,594 | 24 |
| Reboiler | Heat Duty (MWth) | 114.3 | Steam flowrate (t/h) | 14.7 | | | 1,751,281 | 7 |
| Condenser | Cooling Duty (MWth) | 32.2 | Cooling water flowrate (t/h) | 213.7 | | | 693,094 | 3 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 98.7 | Heat transfer area per heat exchanger (m2) | 1500.0 | Number of heat exchangers | 1 | 181,250 | 1 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 391,977 | 2 |
| Lean amine cooler | Cooling Duty (MWth) | 49.4 | Heat transfer area per heat exchanger (m2) | 643.4 | Number of heat exchangers | 4 | 646,069 | 3 |
| Rich solvent pump | Total power requirement (kW) | 532.8 | Number of pumps required | 10 | | | 155,146 | 1 |
| Lean solvent pump | Total power requirement (kW) | 199.9 | Number of pumps required | 5 | | | 76,870 | 0 |
| Cooling water pumps | Total power requirement (kW) | 348.8 | Number of pumps required | 22 | | | 366,790 | 1 |
| Steam boiler | Capacity (t/h steam) | 190.7 | | | | | 1,286,205 | 5 |
| **Total equipment purchase cost (PCE)** | | | | | | | **24,963,780** | |

[1]T/T height – tan to tan height representing the height from the top to the bottom of the column vessel.

# Simulation Results
# Scenario 2 - Capital Expenditure

| Description | £ | % of PCE |
| --- | --- | --- |
| **Equipment purchase cost breakdown** | | |
| Absorber | 26,735,006 | 54 |
| Stripper | 12,095,188 | 24 |
| Reboiler | 3,502,562 | 7 |
| Condenser | 1,386,188 | 3 |
| Lean/Rich Heat Exchanger | 362,501 | 1 |
| Lean amine tank | 783,953 | 2 |
| Lean amine cooler | 1,292,139 | 3 |
| Rich solvent pump | 310,291 | 1 |
| Lean solvent pump | 153,741 | 0 |
| Cooling water pumps | 733,580 | 1 |
| Steam boiler | 2,572,411 | 5 |
| **Total equipment purchase cost (PCE)** | **49,927,561** | |

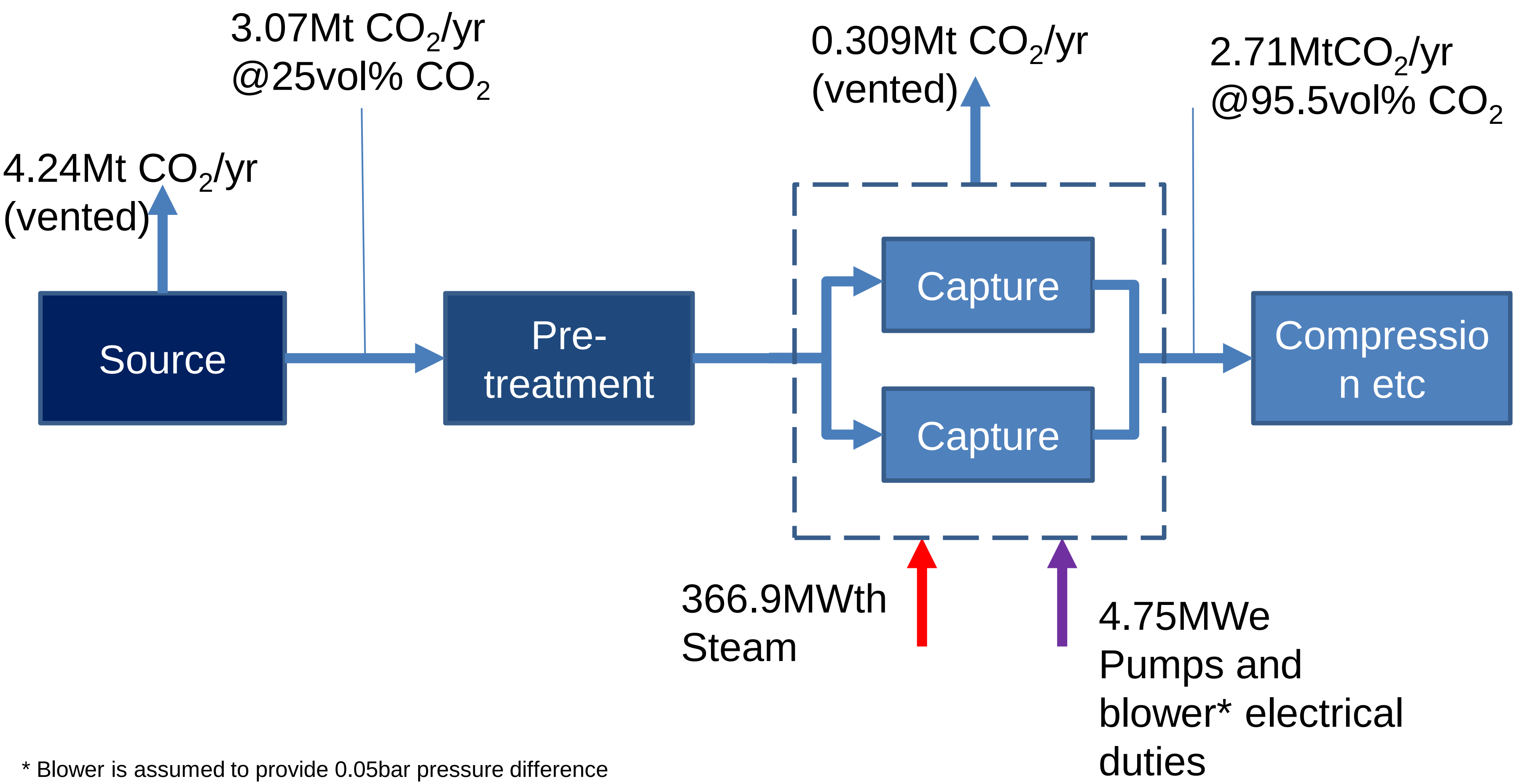
# Simulation Results
# Scenario 2 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 49,927,561 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 159,768,195 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **316,820,331** |

# Simulation Results
# Scenario 2 - Operating Expenditure

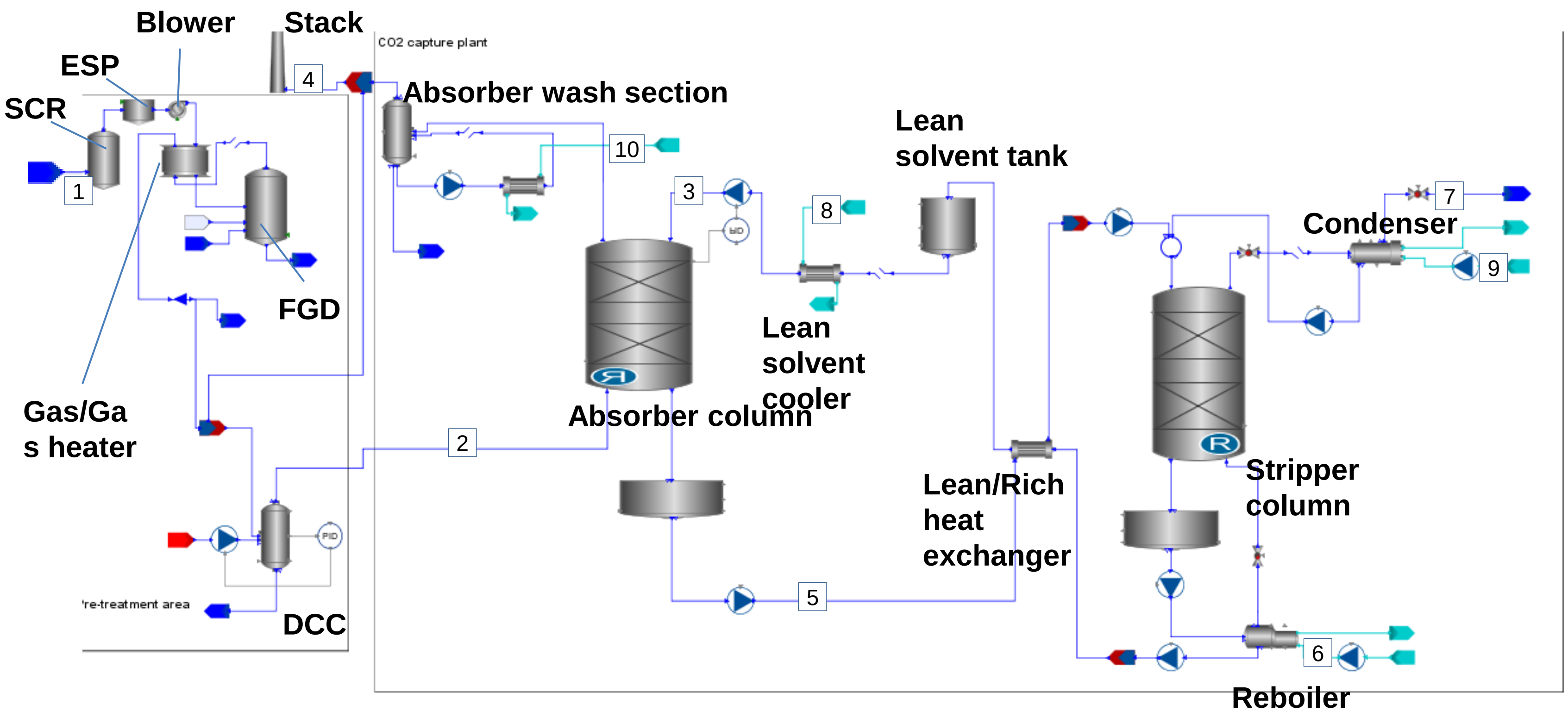
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 15,841,017 |
| Variable Costs | | |
| Solvent make-up costs | 10% of maintenance costs | 1,584,102 |
| Miscellaneous materials cost | | 8,700,300 |
| Pumps power cost | | 1,375,417 |
| Utilities - Steam costs | | 45,550,643 |
| Utilities - Cooling water costs | | 3,261,157 |
| **Total Variable costs** | | **60,471,619** |
| **OPEX** | | **76,312,635** |

## Scenario 3 Process Simulation
## High Level Process/Energy flows (25% $CO_2$)




* Blower is assumed to provide 0.05bar pressure difference

# Scenario 3 Process Simulation
# 25% $CO_2$ in combusted Blast Furnace gas stream




SCR – Selective Catalytic Reduction
ESP – Electrostatic Precipitator
FGD – Flue Gas Desulphurization
DCC – Direct Contact Cooler

# Inputs – Scenario 3

- Two $CO_2$ capture trains
- 30 wt% MEA solvent
- 90% capture target
- >95% purity $CO_2$ by volume
- Absorber operating at atmospheric pressure
- Mellapak 250Y structured packing used in Absorber and stripper columns
- Heat exchanger overall heat transfer coefficient – 6000W/$m^2$K
- Heat exchanger area – 1500$m^2$
- Maximum flow in heat exchanger – 2500$m^3$/hr
- Lean loading specification – 0.225 mol $CO_2$/mol MEA
- Stripper operating pressure ~ 1.67bara
- Steam pressure 3.5bara
- Cooling water temperature is assumed to be 10°C
- 8400 hours of continuous operation assumed in a year

# Stream Tables
# Scenario 3

| From: | | CO2 SOURCE | PRETREATMENT PLANT | LEAN SOLVENT COOLER | ABSORBER | ABSORBER SUMP |
|---|---|---|---|---|---|---|
| **To:** | | PRETREATMENT PLANT | ABSORBER | ABSORBER | STACK | LEAN/RICH HEAT EXCHANGER |
| **Service:** | | FLUE GAS | FLUE GAS* | LEAN AMINE* | OUTLET GAS | RICH AMINE* |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | VAPOUR | LIQUID |
| **Stream Number:** | | 1 | 2 | 3 | 4 | 5 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 47,969 | 22,965 | 1,843,732 | 30,937 | 1,746,725 |
| **MEA** | | 0 | 0.0 | 849159.4 | 0.0 | 848415.4 |
| **$CO_2$** | | 365,787 | 182,924 | 137,640 | 36,634 | 299,184 |
| **$N_2$** | | 602,234 | 314,393 | 0 | 616,211 | 53 |
| **$O_2$** | | 24,520 | 0 | 0 | 12,452 | 0 |
| **$SO_2$** | | 262 | 0 | 0 | 4 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 1,040,904 | 520,282 | 2,830,531 | 696,248 | 2,894,377 |
| **Temperature** | ºC | 60.0 | 40.9 | 40.8 | 40.0 | 59.0 |
| **Pressure** | bar(a) | 1.01 | 1.10 | 1.07 | 1.07 | 1.10 |

* This value corresponds to a single train

# Stream Tables (continued)
## Scenario 3

| | | | | | | |
|---|---|---|---|---|---|---|
| **From:** | | STEAM SUPPLY | STRIPPER CONDENSER | COOLING WATER SUPPLY | COOLING WATER SUPPLY | COOLING WATER SUPPLY |
| **To:** | | REBOILER | COMPRESSION | LEAN AMINE COOLER | STRIPPER CONDENSER | ABSORBER WASH SECTION COOLER |
| **Service:** | | STEAM | CO2 PRODUCT | COOLING WATER | COOLING WATER | COOLING WATER |
| **Phase:** | | VAPOUR | VAPOUR | LIQUID | LIQUID | LIQUID |
| **Stream Number:** | | 6 | 7 | 8 | 9 | 10 |
| **Mass Flow** | kg/hr | | | | | |
| **$H_2O$** | | 612,173 | 6,099 | 8,282,783 | 7,989,153 | 23,441,026 |
| **MEA** | | 0.0 | 3.0 | 0.0 | 0.0 | 0.0 |
| **$CO_2$** | | 0 | 323,006 | 0 | 0 | 0 |
| **$N_2$** | | 0 | 105 | 0 | 0 | 0 |
| **$O_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **$SO_3$** | | 0 | 0 | 0 | 0 | 0 |
| **$NO_2$** | | 0 | 0 | 0 | 0 | 0 |
| **CO** | | 0 | 0 | 0 | 0 | 0 |
| **Particulates** | | 0 | 0 | 0 | 0 | 0 |
| | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 612,173 | 329,213 | 8,282,783 | 7,989,153 | 23,441,026 |
| **Temperature** | ºC | 127.5 | 39.8 | 9.9 | 25.0 | 9.9 |
| **Pressure** | bar(a) | 3.1 | 1.5 | 1.0 | 2.0 | 1.0 |

# Simulation Results
# Scenario 3 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of capture plant | 2 |
| Source % $CO_2$ | 25 |
| Site total $CO_2$ captureable (tonnes/year) | 7,305,903 |
| % site $CO_2$ captureable | 43 |
| Total Reboiler Heat duty (MWth) | 366.9 |
| Reboiler Specific duty (GJ/t $CO_2$) | 4.09 |
| Lean loading (mol $CO_2$/mol MEA) | 0.225 |
| Total electrical power requirement of capture plant pumps (MWe) | 3.11 |
| Electrical power requirement of blower* (MWe) | 1.49 |
| Cooling water required (tonnes/hr) | 35,158 |
| Capture plant site area required ($m^2$) | 25,156 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 95.5<br>$H_2O$ – 4.4<br>$N_2$ – 0.08 |
| Non-$CO_2$ emissions to atmosphere | |
| Before (ppm) | NOx – 100<br>SOx – 123 |
| After (ppm) | NOx – 6<br>SOx – 2 |

* Blower is assumed to raise the pressure of flue gas by 0.05bar

# Simulation Results
# Scenario 3 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Absorber | Diameter (m) | 11.5 | Packing Height (m) | 21.3 | T/T Height (m) | 51.3 | 14,662,105 | 48 |
| Stripper | Diameter (m) | 10.3 | Packing Height (m) | 10.0 | T/T Height (m) | 40 | 7,150,601 | 23 |
| Reboiler | Heat Duty (MWth) | 183.4 | Steam flowrate (t/h) | 23.6 | | | 3,013,777 | 10 |
| Condenser | Cooling Duty (MWth) | 46.4 | Cooling water flowrate (t/h) | 308.2 | | | 1,050,394 | 3 |
| Lean/Rich Heat Exchanger | Heat Duty (MWth) | 122.3 | Heat transfer area per heat exchanger (m2) | 750.0 | Number of heat exchangers | 2 | 377,826 | 1 |
| Lean amine tank | Volume of tank (m3) | 785.4 | | | | | 483,367 | 2 |
| Lean amine cooler | Cooling Duty (MWth) | 96.3 | Heat transfer area per heat exchanger (m2) | 759.2 | Number of heat exchangers | 6 | 1,147,221 | 4 |
| Rich solvent pump | Total power requirement (kW) | 791.7 | Number of pumps required | 14 | | | 224,582 | 1 |
| Lean solvent pump | Total power requirement (kW) | 294.8 | Number of pumps required | 7 | | | 110,764 | 0 |
| Cooling water pumps | Total power requirement (kW) | 466.6 | Number of pumps required | 36 | | | 606,178 | 2 |
| Steam boiler | Capacity (t/h steam) | 306.1 | | | | | 2,017,216 | 7 |
| **Total equipment purchase cost (PCE)** | | | | | | | **30,844,032** | |

[1]T/T height – tan to tan height representing the height from the top to the bottom of the column vessel.

# Simulation Results
# Scenario 3 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Absorber | 29,324,210 | 47.5 |
| Stripper | 14,301,202 | 23.2 |
| Reboiler | 6,027,554 | 9.8 |
| Condenser | 2,100,789 | 3.4 |
| Lean/Rich Heat Exchanger | 755,653 | 1.2 |
| Lean amine tank | 966,733 | 1.6 |
| Lean amine cooler | 2,294,443 | 3.7 |
| Rich solvent pump | 449,164 | 0.7 |
| Lean solvent pump | 221,528 | 0.4 |
| Cooling water pumps | 1,212,356 | 2.0 |
| Steam boiler | 4,034,431 | 6.5 |
| Total equipment purchase cost (PCE) | **61,688,063** | |

# Simulation Results
## Scenario 3 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 61,688,063 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 197,401,803 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Solvent initial Charge | 5 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | **391,447,775** |

# Simulation Results
# Scenario 3 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 19,572,389 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 1,957,239 |
| Solvent make-up cost | | 13,689,255 |
| Pumps power cost | | 1,975,296 |
| Utilities - Steam costs | | 73,122,804 |
| Utilities - Cooling water costs | | 4,429,848 |
| **Total Variable costs** | | **95,174,442** |
| **OPEX** | | **114,746,831** |

# Comparison of costs between scenarios

| | Baseline Scenario #1 | Sensitivity "15% CO2 from Blast Furnace" (Scenario #2) | Sensitivity "25% CO2 from Blast Furnace" (Scenario #3) |
|---|---|---|---|
| Source $CO_2$ | 2.53 Mt/yr | 1.95 Mt/yr | 3.07Mt/yr |
| Equipment cost | £55m | £50m | £62m |
| Total fixed cost | £294m | £267m | £330m |
| Annual opex (incl. energy) | £96m/yr | £76m/yr | £115m/yr |
| Reboiler Heat Duty $MW_{th}$ | 299 | 229 | 367 |
| Power /$MW_e$ | 4.21 | 3.73 | 4.75 |

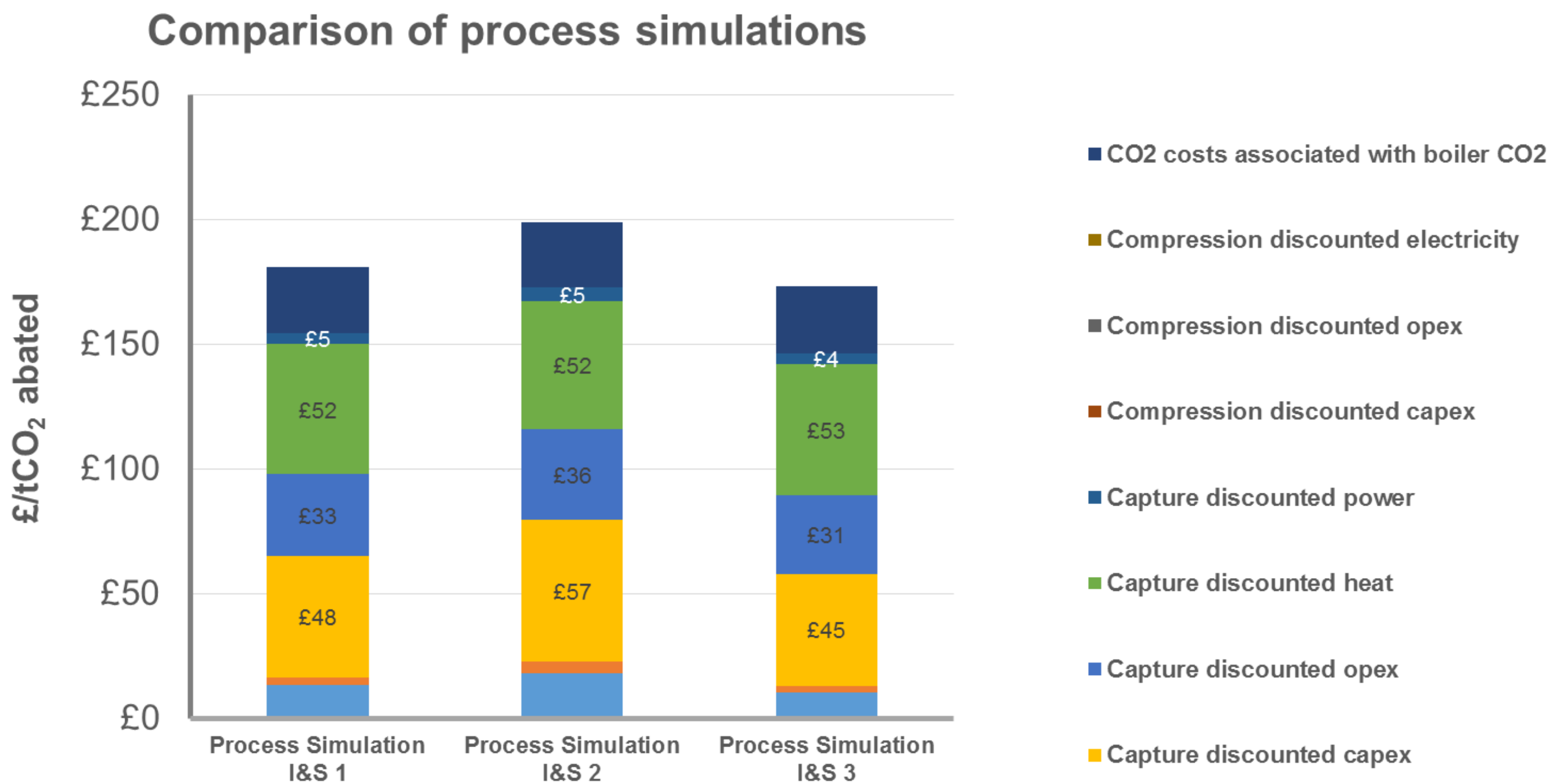
# Insights from process simulation

- Capture with a 1st generation amine solvent at an iron and steel site is feasible, but there are many potential alternative configurations that should be considered.
- The overall capital cost is *ca*. 6 times the cost of the main pieces of equipment.
- The largest of the equipment cost items modelled is the absorber.
- The largest operating cost modelled is for steam.
- Therefore capture technology development should focus on reducing the costs of absorber, and/or the amount of steam required, and simplifying retrofit installation.
- Given a scale of project will adopt configurations involving two absorber trains rather than one.
- Stakeholders should review water requirements with appropriate stakeholders.

# The outputs from the process simulation and techno-economic model can be compared to prioritise future model refinements

| Parameter | Techno-economics "Baseline" | Process simulation "Baseline" |
| --- | --- | --- |
| Input flue gas $MtCO_2$/yr | 3.9 Mt/yr captured out of total site 7.3 Mt/yr | 2.5 Mt/yr |
| Abated $MtCO_2$/yr | 3.2 $MtCO_2$/yr | Not calculated directly |
| Capex | Capture only: £157m (£216m incl. pre-treatment) | Capture only: £349m |
| Non-energy opex | £10m/yr | £32m/yr |
| Heat | 94 MW | 299 MW |
| Power | 2.2 MW | 4.2 MW |

# Comparison of the abatement cost breakdown for three process simulations

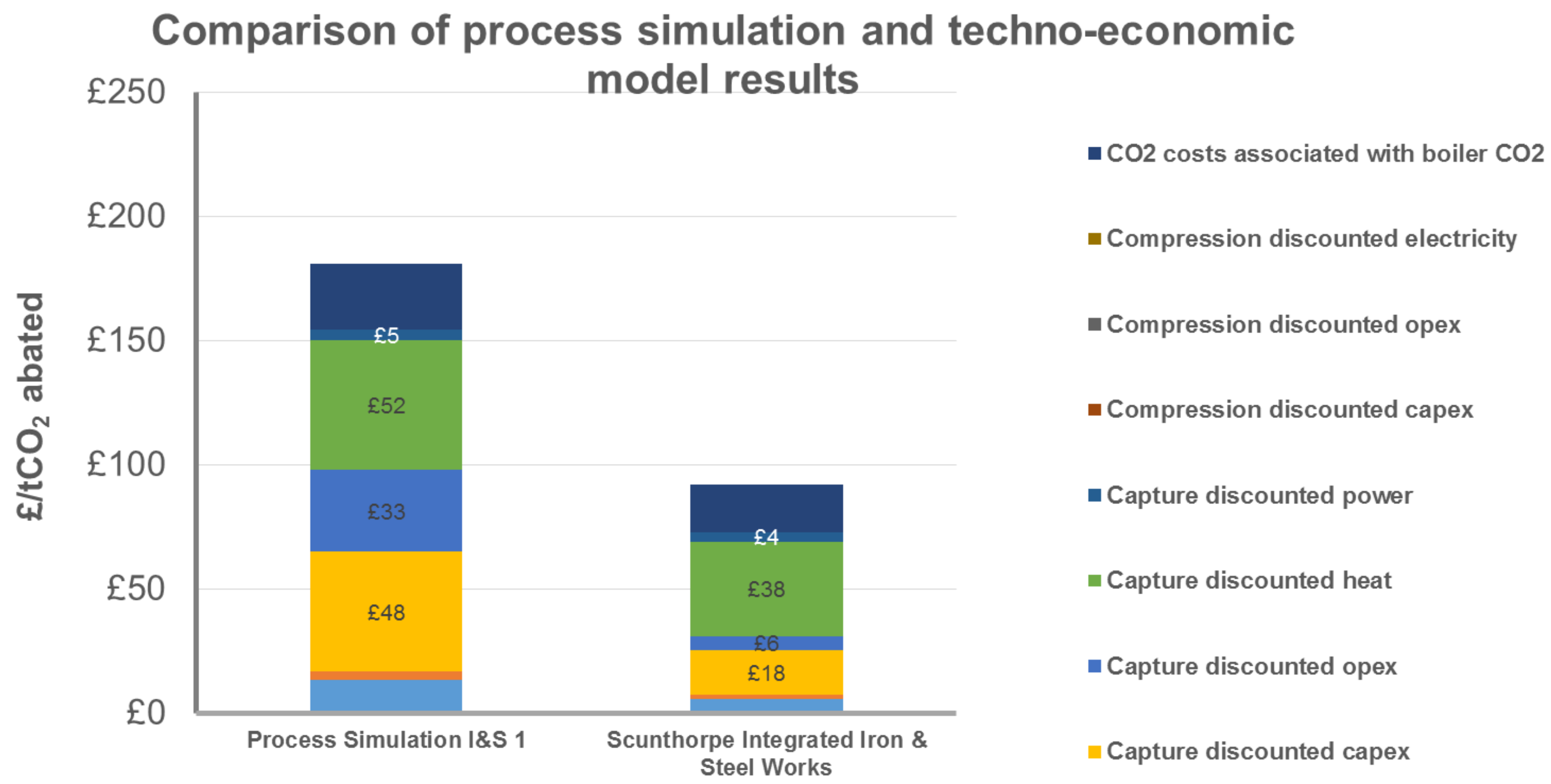



20% $CO_2$
2.5 Mt/yr input
1 Mt/yr abated

15% $CO_2$
1 Mt/yr input
0.8 Mt/yr abated

25% $CO_2$
3 Mt/yr input
1.3 Mt/yr abated

# Comparison of baseline abatement cost breakdown for the process simulation and techno-economic modelling for Iron and Steel sector




2.5 Mt/yr
input
20% $CO_2$
1 Mt/yr
abated

30% $CO_2$
3.2 Mt/yr
abated

# Four case studies on process simulation

1. Grangemouth oil refinery with MEA capture
2. Lafarge Tarmac Dunbar cement works with MEA capture
3. Tata Scunthorpe iron and steel plant with MEA capture at CHP unit
4. GrowHow ammonia production site with $CO_2$ compression

# Scope of study

- The GrowHow Teesside kindly offered to provide data on $CO_2$ streams for process simulation, as an example of a chemical sector site with realistic capture potential in the period to 2025 (or earlier).
- The $CO_2$ stream is high purity (>99%), and therefore the primary requirement is likely to be for compression (and dehydration), and it is assumed that no $CO_2$ capture is required.
- Key uncertainties identified and prioritised for process simulation of compression (not just at GrowHow) include capacity (Mt/yr), output $CO_2$ pressure (i.e. gas phase or dense phase), and requirements for dehydration for wet $CO_2$.
- Therefore six scenarios were developed to examine these issues.

| Scenario | Scale (Mt/yr) | Output pressure |
|---|---|---|
| 1 | 0.5 | 40 bar |
| 2 | 0.1 | 40 bar |
| 3 | 2 | 40 bar |
| 4 | 0.5 | 110 bar |
| 5 | 2 | 110 bar |
| 6 (95% pure $CO_2$) | 0.5 | 110 bar |

# Detailed assumptions

- Dehydrator capital expenditure includes cost of fired heater for bed regeneration
- Electric driver costs were extrapolated beyond the stated range of the available cost function
- Footprint of compression train was estimated based on published work[1] and scaled based on the number of compressor sections required.
- Knock-out drums were sized based on settling velocities[2]
- Corrosion margin in knock-out drums assumed to be 2mm
- Distance of cooling water supply was not taken into account
- Cooling water temperature is available at 25°C. Cooling water return temperature is 35°C
- Number of compressor sections is selected to avoid compressor discharge temperatures of more than 150°C. In between compressor sections, the CO2 can be cooled in coolers (heat exchangers).

[1] IEAGHG (2012). $CO_2$ capture at gas-fired power plants, Report No: 2012/8.
[2] GPSA Engineering Data Book (2004).

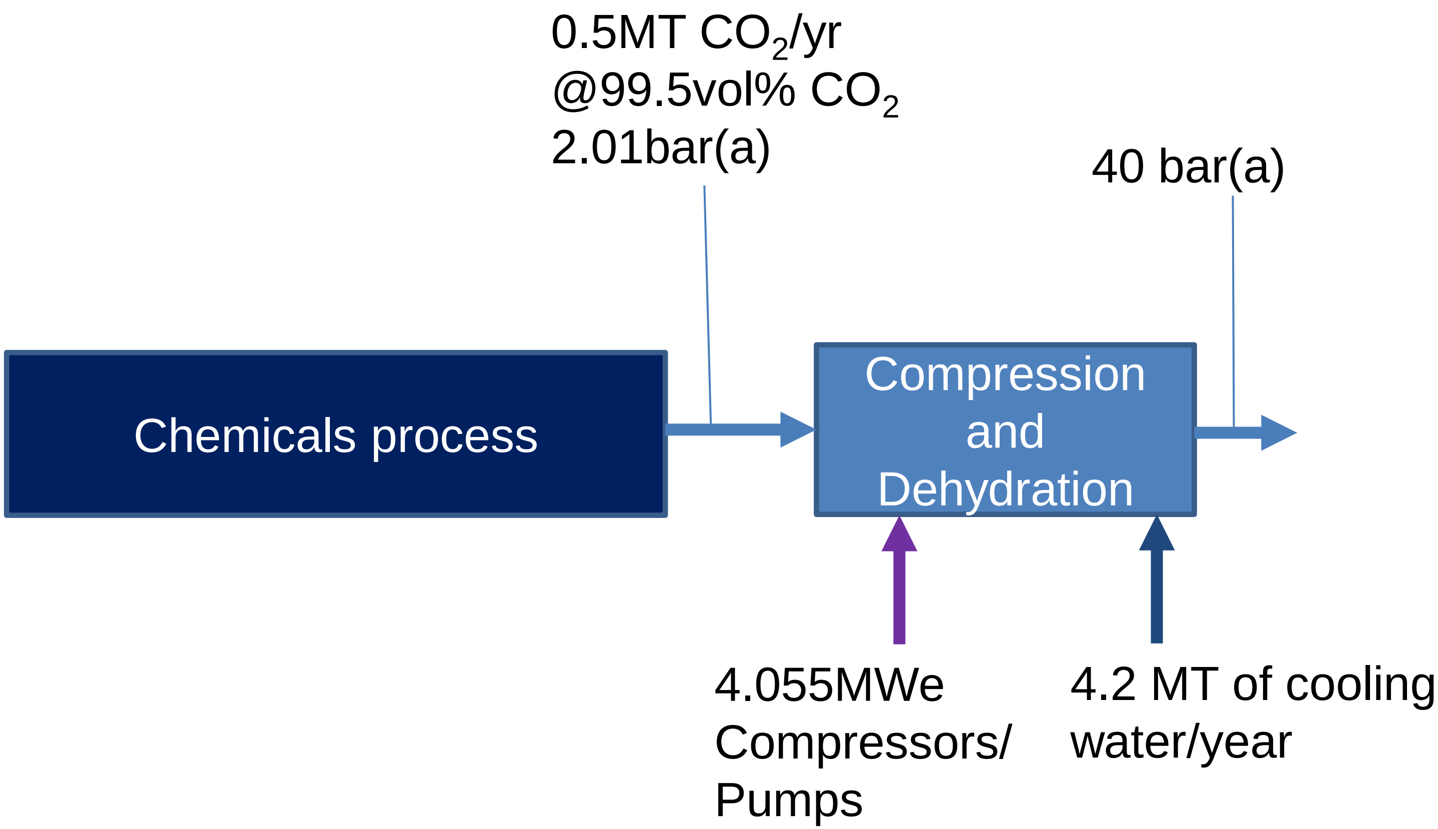
# Detailed assumptions

- A constant compressor shaft speed of 80Hz was assumed
- 8400 hours of continuous operation assumed in a year
- Stainless steel used is 304L
- No limitations to cooling water or electricity supply were considered
- Number of compressor trains were sized based on technical feasibility (based on maximum impeller diameters)
- Each compressor frame could have one or two frames. Two frames are selected if compression is required after the dehydrator
- Double pipe heat exchangers are used where heat transfer areas required are small ($<80m^2$)
- Electricity costs are based on DECC's 2025 prices for the demo scale scenarios and 2020 prices for the pilot plant scenario.
- All other costs are based on Q3 2013. These could be translated to the same basis by assuming an appropriate discount rate.

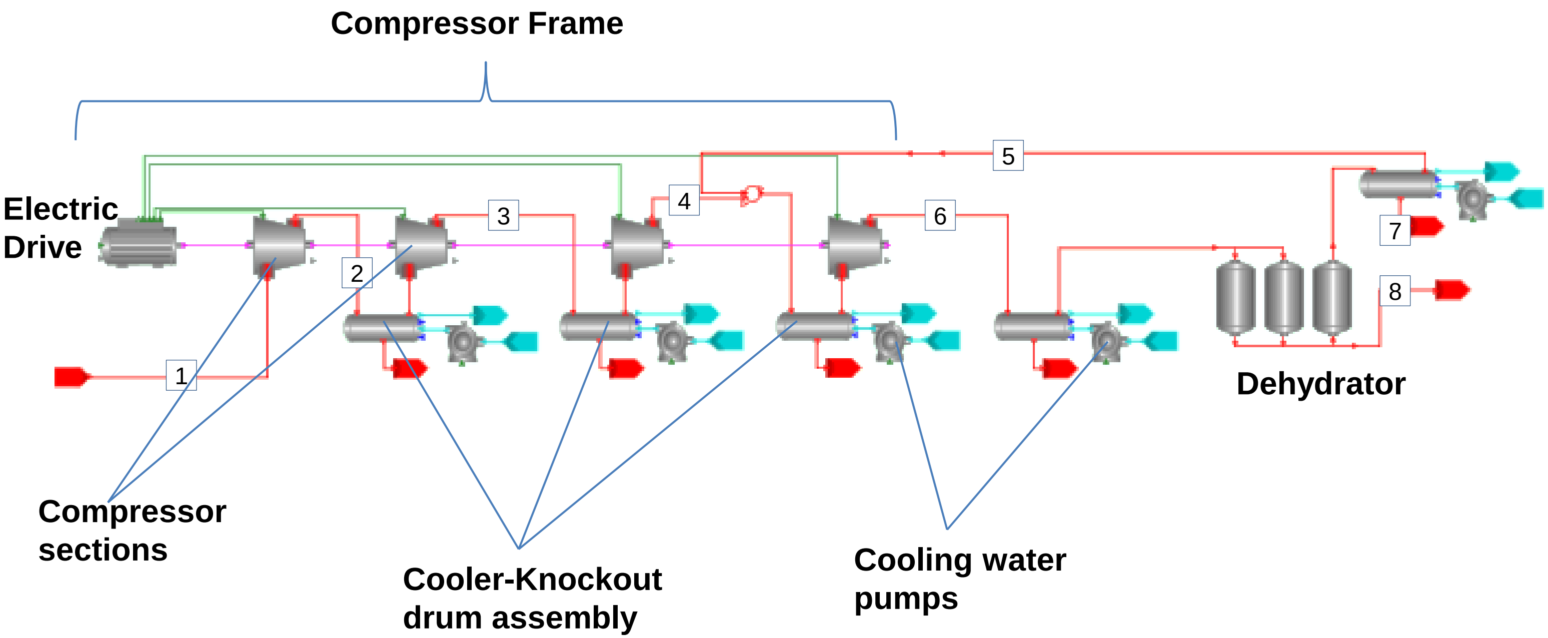
# Narrative

Carbon dioxide produced in the ammonia plant at high purity levels and at about 1 bar(g) pressure. This fluid is compressed through a series of compressor sections with cooling in between sections.  A dehydration unit reduces moisture content to 50ppmv.

## Baseline Process Simulation
## High Level Process/Energy flows



# Baseline Process Simulation



# Inputs - Baseline

- One $CO_2$ compressor train
- 0.5MT $CO_2$/year
- 99.5 vol% $CO_2$ at inlet
- 1bar(g) inlet pressure
- 40bar(a) discharge pressure
- 30°C inlet temperature
- Interstage cooling target - 40°C
- 80Hz compressor frame speed
- Dehydrator moisture specification – 50ppm
- Cooling water temperature is assumed to be 25°C

# Stream Tables
# Baseline scenario

| | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| From: | | CO2 SOURCE | FIRST COMPRESSOR | SECOND COMPRESSOR | THIRD COMPRESSOR | RECYCLE COOLERKODRUM | FOURTH COMPRESSOR | RECYCLE COOLERKODRUM | DEHYDRATOR |
| To: | | FIRST COMPRESSOR | FIRST COOLERKODRUM | SECOND COOLERKODRUM | RECYCLE MIXER | RECYCLE MIXER | FOURTH COOLERKODRUM | WASTE WATER | OUTLET CO2 |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | BOTTOMS | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | LIQUID | VAPOUR |
| Stream Number: | | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
| **Mass Flow** | kg/hr | | | | | | | | |
| $CO_2$ | | 59,757 | 59,757 | 59,757 | 59,757 | 2,232 | 61,989 | 0 | 59,757 |
| $CH_4$ | | 1 | 1 | 1 | 1 | 0 | 1 | 0 | 1 |
| CO | | 2 | 2 | 2 | 2 | 0 | 2 | 0 | 2 |
| $H_2O$ | | 25 | 25 | 25 | 25 | 3.5 | 28.1 | 23.4 | 1 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 52 | 52 | 52 | 52 | 2 | 54 | 0 | 52 |
| $H_2$ | | 8 | 8 | 8 | 8 | 0 | 8 | 0 | 8 |
| $CH_3OH$ | | 1 | 1 | 1 | 1 | 0 | 1 | 0 | 1 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 59,846 | 59,846 | 59,846 | 59,846 | 2,237 | 62,083 | 23 | 59,822 |
| Temperature | ºC | 30.0 | 119.6 | 116.4 | 100.8 | 40.0 | 93.3 | 40.0 | 40.4 |
| Pressure | Bar (a) | 2.01 | 5.61 | 12.51 | 23.68 | 23.68 | 41.2 | 23.7 | 40.0 |

# Simulation Results
## Baseline scenario - Process Conditions

| Description | Value |
|---|---|
| Number of trains of compression | 1 |
| Number of compressor frames | 1 |
| Number of compressor sections | 4 |
| Source vol % $CO_2$ | 99.5 |
| $CO_2$ for transportation (tonnes/year) | 501,961 |
| Total compressor electrical power requirement (MWe) | 4.04 |
| Total electrical power requirement of compression cooling water pumps (kWe) | 17 |
| Total Cooling water required (tonnes/hr) | 501.1 |
| Total Capture plant site area required ($m^2$) | 2600 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 99.6 |
| | $H_2O$ – 0.005 |
| | $N_2$ – 0.14 |
| | $H_2$ – 0.29 |

# Simulation Results
## Baseline scenario – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Compressor frame | Electrical Duty (MWe) | 4.0 | | | | | 2,180,860 | 74.2 |
| Heat exchanger | Range of heat exchanger heat transfer area ($m^2$) | 93-107 | Number of shell and tube exchangers/double pipe exchangers | 4/1 | Operating pressure ranges bar(a) | 5-41 | 130,612 | 4.4 |
| Knockout drums | Number of knockout drums | 5 | Range of mass of steel required for construction (kg) | 120-1418 | | | 146,991 | 5.0 |
| Electric Drives | Electrical Duty (MWe) | 4.0 | | | | | 106,046 | 3.6 |
| Dehydrator | Water capacity (kg/hr water adsorbed) | 26.90 | | | | | 332,442 | 11.3 |
| Pumps | Electrical Duty (kWe) | 16.9 | | | | | 40,378 | 1.4 |
| Total equipment purchase cost (PCE) | | | | | | | 2,937,327 | |

# Baseline scenario - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Compressor frame | 2,180,860 | 74.2 |
| Heat exchanger | 130,612 | 4.4 |
| Knockout drums | 146,991 | 5.0 |
| Electric Drives | 106,046 | 3.6 |
| Dehydrator | 332,442 | 11.3 |
| Pumps | 40,378 | 1.4 |
| Total equipment purchase cost (PCE) | 2,937,327 | |

# Simulation Results
## Baseline scenario - Capital Expenditure

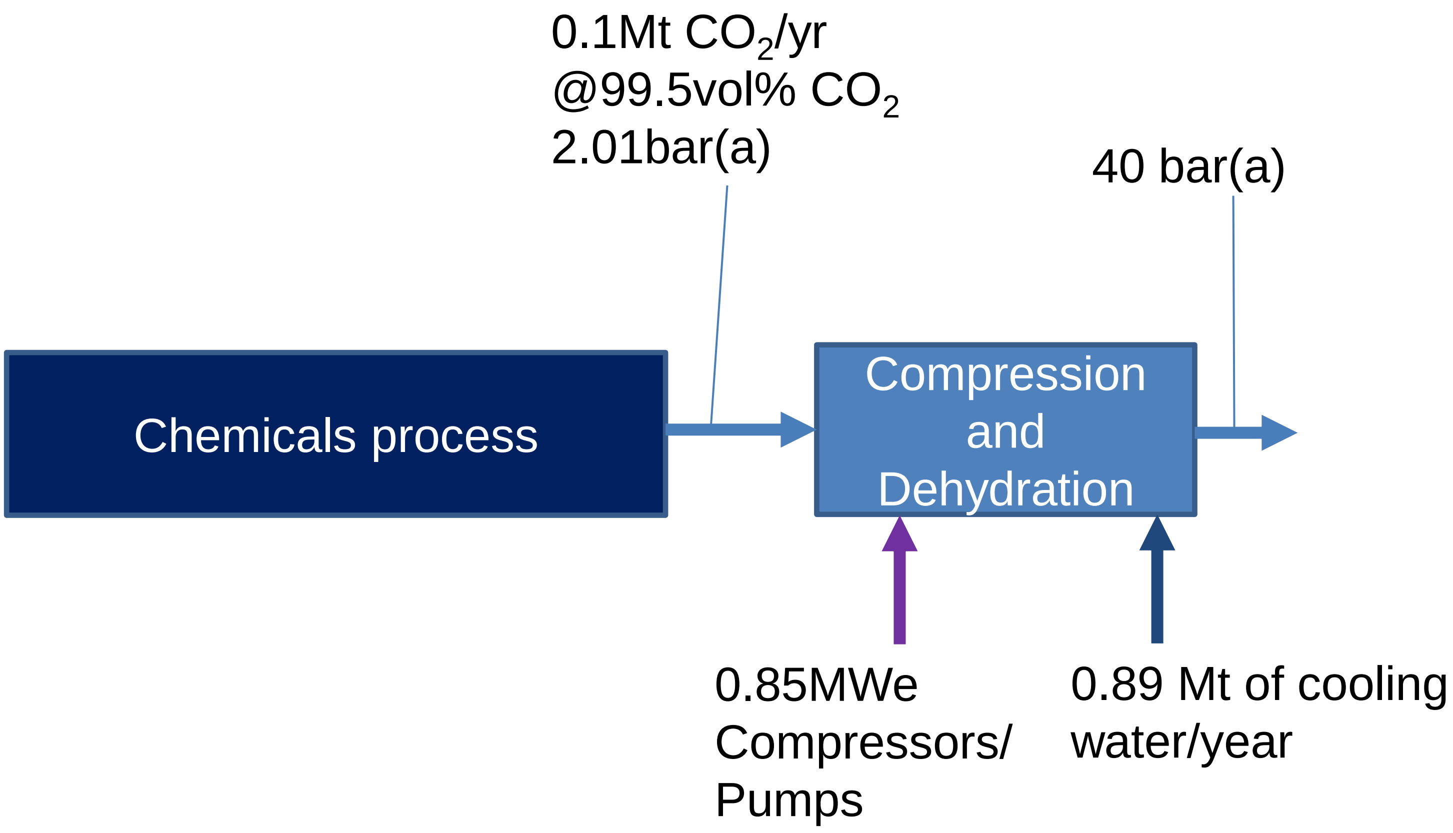
| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 2,937,327 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 9,399,448 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | 18,169,132 |

# Simulation Results
## Baseline scenario - Operating Expenditure

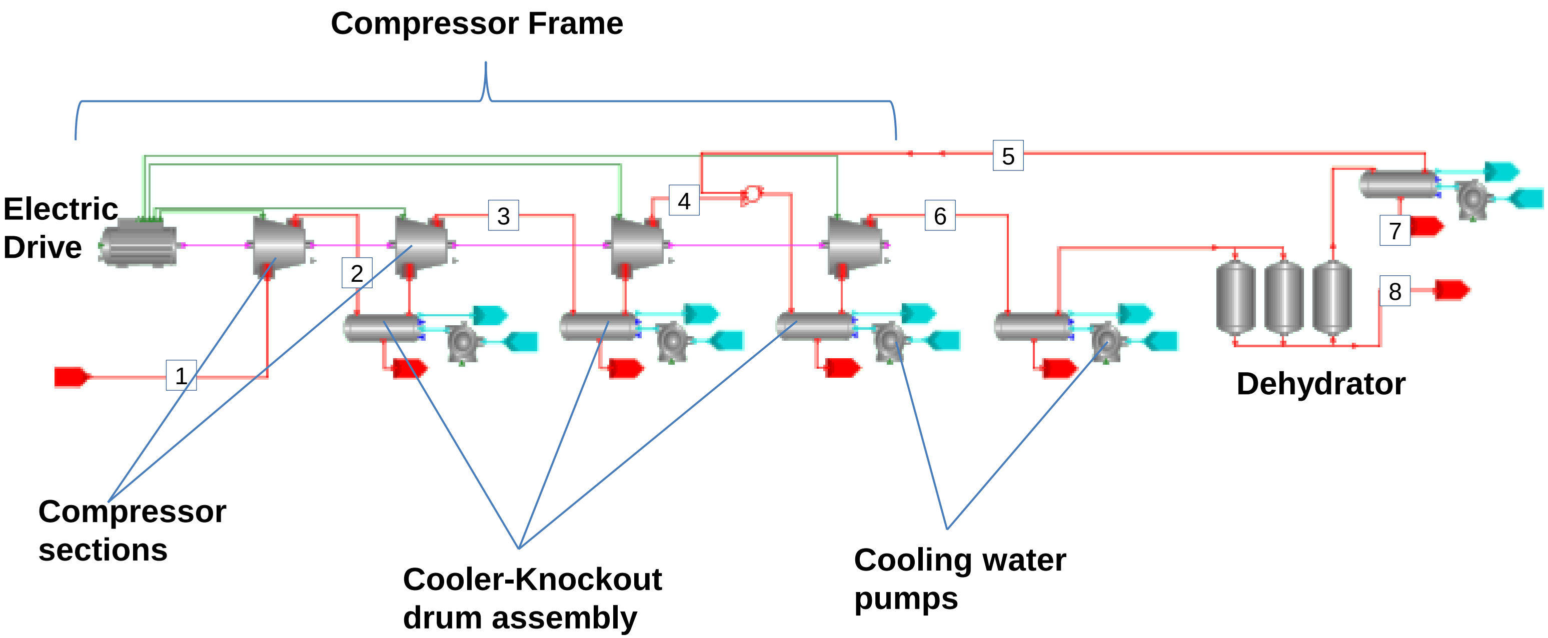
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 908,457 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 90,846 |
| Compressor electrical power cost | | 3,790,398 |
| Cooling water pumps power cost | | 10,368 |
| Dehydrator heating costs | | 31,919 |
| Utilities - Cooling water costs | | 96,365 |
| Total Variable costs | | 4,019,895 |
| OPEX | | 4,928,351 |

## Scenario 2 Process Simulation

# Gas phase, pilot (0.1Mt $CO_2$/year)



# Scenario 2 Process Simulation



# Inputs – Scenario 2

- One $CO_2$ compressor train
- 0.1MT $CO_2$/year
- 99.5 vol% $CO_2$ at inlet
- 1bar(g) inlet pressure
- 40bar(a) discharge pressure
- 30°C inlet temperature
- Interstage cooling target - 40°C
- 80Hz compressor frame speed
- Dehydrator moisture specification – 50ppm
- Cooling water temperature is assumed to be 25°C

# Stream Tables
## Scenario 2

| From: | | CO2 SOURCE | FIRST COMPRESSOR | SECOND COMPRESSOR | THIRD COMPRESSOR | RECYCLE COOLERKODRUM | FOURTH COMPRESSOR | RECYCLE COOLERKODRUM | DEHYDRATOR |
|---|---|---|---|---|---|---|---|---|---|
| To: | | FIRST COMPRESSOR | FIRST COOLERKODRUM | SECOND COOLERKODRUM | RECYCLE MIXER | RECYCLE MIXER | FOURTH COOLERKODRUM | WASTE WATER | OUTLET CO2 |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | BOTTOMS | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | LIQUID | VAPOUR |
| Stream Number: | | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
| **Mass Flow** | kg/hr | | | | | | | | |
| $CO_2$ | | 11,952 | 11,952 | 11,952 | 11,952 | 446 | 12,398 | 0 | 11,952 |
| $CH_4$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| CO | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $H_2O$ | | 5 | 5 | 5 | 5 | 0.7 | 5.6 | 4.3 | 1 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 10 | 10 | 10 | 10 | 0 | 11 | 0 | 10 |
| $H_2$ | | 2 | 2 | 2 | 2 | 0 | 2 | 0 | 2 |
| $CH_3OH$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 11,969 | 11,969 | 11,969 | 11,969 | 448 | 12,417 | 4 | 11,965 |
| Temperature | ºC | 30.0 | 123.7 | 119.4 | 103.6 | 40.0 | 97.0 | 40.0 | 40.3 |
| Pressure | Bar (a) | 2.01 | 5.61 | 12.51 | 23.68 | 23.68 | 41.2 | 23.7 | 40.0 |

# Scenario 2 - Process Conditions

| Description | Value |
| --- | --- |
| Number of trains of compression | 1 |
| Number of compressor frames | 1 |
| Number of compressor sections | 4 |
| Source vol % $CO_2$ | 99.5 |
| $CO_2$ for transportation (tonnes/year) | 100,393 |
| Total compressor electrical power requirement (MWe) | 0.85 |
| Total electrical power requirement of compression cooling water pumps (kWe) | 4 |
| Total Cooling water required (tonnes/hr) | 105.4 |
| Total Capture plant site area required ($m^2$) | 2600 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 99.6 |
| | $H_2O$ – 0.005 |
| | $N_2$ – 0.14 |
| | $H_2$ – 0.29 |

# Simulation Results

# Scenario 2 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Compressor frame | Electrical Duty (MWe) | 0.9 | | | | | 1,081,618 | 72.7 |
| Heat exchanger | Range of heat exchanger heat transfer area (m$^2$) | 0.6-11 | Number of shell and tube exchangers/double pipe exchangers | 0/5 | Operating pressure ranges bar(a) | 5-41 | 84,251 | 5.7 |
| Knockout drums | Number of knockout drums | 5 | Range of mass of steel required for construction (kg) | 120-218 | | | 74,091 | 5.0 |
| Electric Drives | Electrical Duty (MWe) | 0.9 | | | | | 66,991 | 4.5 |
| Dehydrator | Water capacity (kg/hr water adsorbed) | 5.03 | | | | | 151,864 | 10.2 |
| Pumps | Electrical Duty (kWe) | 3.5 | | | | | 29,030 | 2.0 |
| Total equipment purchase cost (PCE) | | | | | | | 1,487,844 | |

# Scenario 2 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Compressor frame | 1,081,618 | 72.7 |
| Heat exchanger | 84,251 | 5.7 |
| Knockout drums | 74,091 | 5.0 |
| Electric Drives | 66,991 | 4.5 |
| Dehydrator | 151,864 | 10.2 |
| Pumps | 29,030 | 2.0 |
| **Total equipment purchase cost (PCE)** | 1,487,844 | |

# Scenario 2 - Capital Expenditure

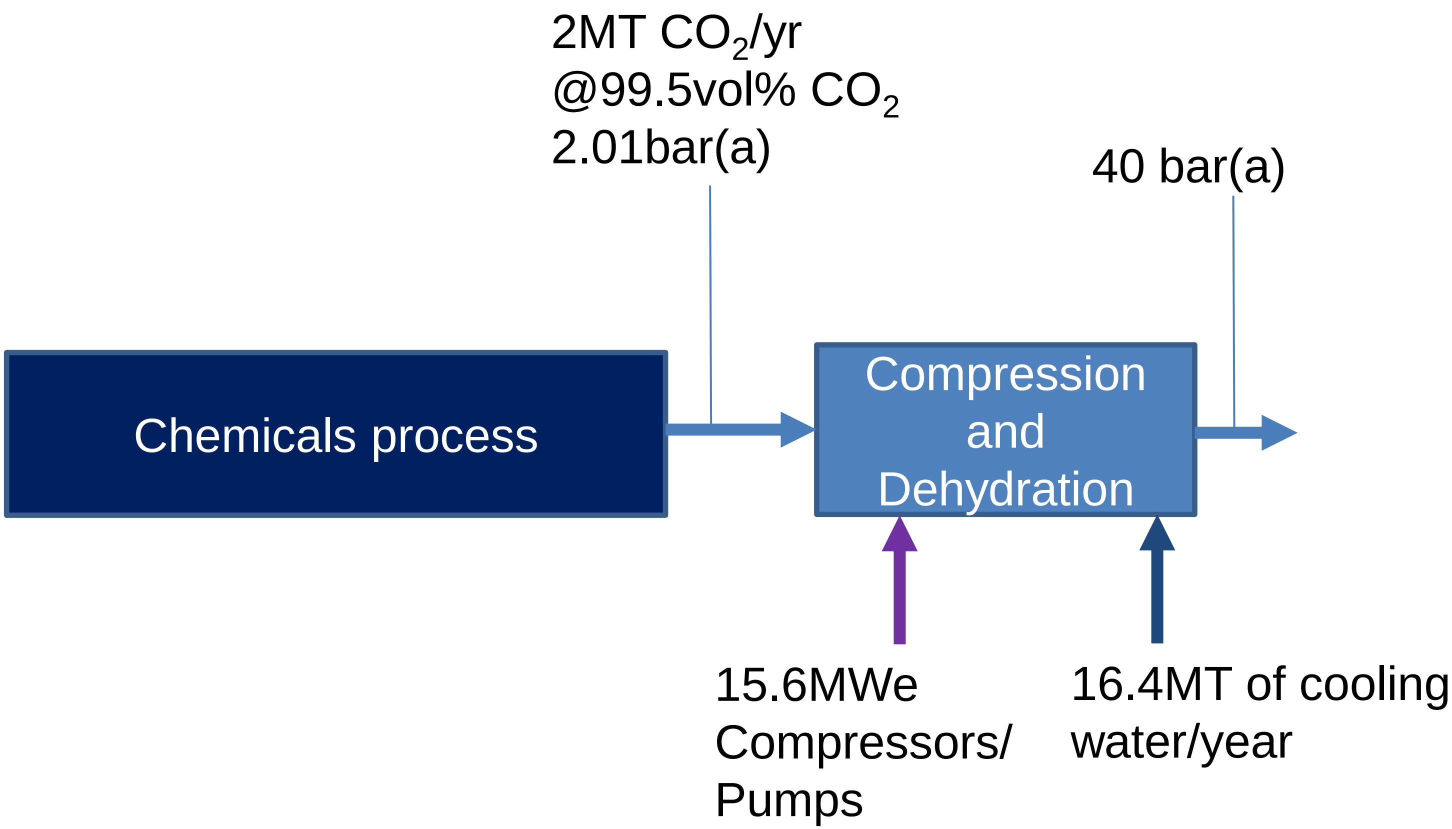
| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 1,487,844 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 4,761,099 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | 9,203,205 |

# Simulation Results

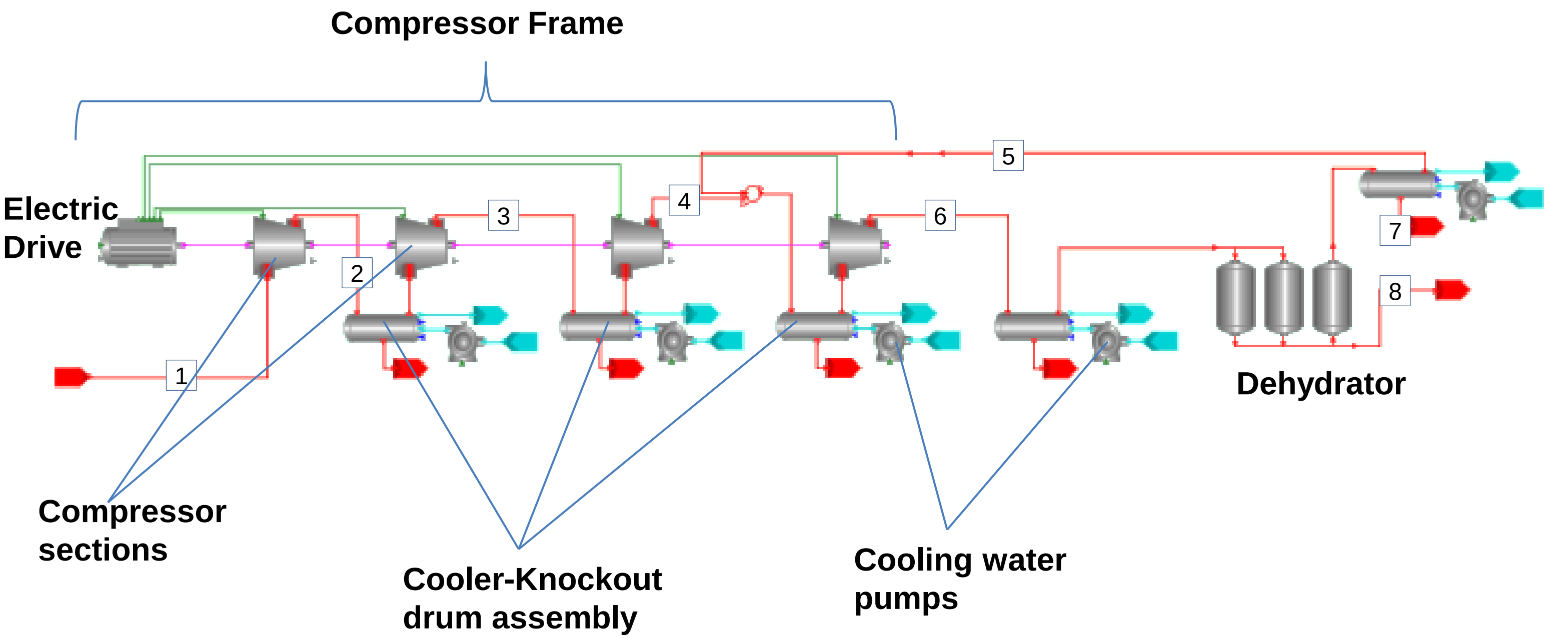
# Scenario 2 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 460,160 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 46,016 |
| Compressor electrical power cost | | 525,565 |
| Cooling water pumps power cost | | 2,180 |
| Dehydrator heating costs | | 6,387 |
| Utilities - Cooling water costs | | 20,263 |
| **Total Variable costs** | | 600,411 |
| **OPEX** | | 1,060,571 |

# Gas phase, large (2MT CO2/year)



# Scenario 3 Process Simulation



# Inputs – Scenario 3

- One $CO_2$ compressor train
- 2MT $CO_2$/year
- 99.5 vol% $CO_2$ at inlet
- 1bar(g) inlet pressure
- 40bar(a) discharge pressure
- 30°C inlet temperature
- Interstage cooling target – 40°C
- 80Hz compressor frame speed
- Dehydrator moisture specification – 50ppm
- Cooling water temperature is assumed to be 25°C

# Stream Tables
## Scenario 3

| From: | | CO2 SOURCE | FIRST COMPRESSOR | SECOND COMPRESSOR | THIRD COMPRESSOR | RECYCLE COOLERKODRUM | FOURTH COMPRESSOR | RECYCLE COOLERKODRUM | DEHYDRATOR |
|---|---|---|---|---|---|---|---|---|---|
| To: | | FIRST COMPRESSOR | FIRST COOLERKODRUM | SECOND COOLERKODRUM | RECYCLE MIXER | RECYCLE MIXER | FOURTH COOLERKODRUM | WASTE WATER | OUTLET CO2 |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | BOTTOMS | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | LIQUID | VAPOUR |
| Stream Number: | | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 |
| **Mass Flow** | kg/hr | | | | | | | | |
| $CO_2$ | | 239,029 | 239,029 | 239,029 | 239,029 | 8,926 | 247,955 | 0 | 239,029 |
| $CH_4$ | | 4 | 4 | 4 | 4 | 0 | 5 | 0 | 4 |
| CO | | 8 | 8 | 8 | 8 | 0 | 8 | 0 | 8 |
| $H_2O$ | | 98 | 98 | 98 | 98 | 14.1 | 112.5 | 86.4 | 12 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 206 | 206 | 206 | 206 | 8 | 214 | 0 | 206 |
| $H_2$ | | 32 | 32 | 32 | 32 | 1 | 33 | 0 | 32 |
| $CH_3OH$ | | 4 | 4 | 4 | 4 | 0 | 4 | 0 | 4 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 239,382 | 239,382 | 239,382 | 239,382 | 8,950 | 248,332 | 87 | 239,296 |
| Temperature | ºC | 30.0 | 117.2 | 113.7 | 98.9 | 40.0 | 92.0 | 40.0 | 40.3 |
| Pressure | bar(a) | 2.01 | 5.61 | 12.51 | 23.68 | 23.68 | 41.2 | 23.7 | 40.0 |

## Simulation Results

# Scenario 3 - Process Conditions

| Description | Value |
| --- | --- |
| Number of trains of compression | 1 |
| Number of compressor frames | 1 |
| Number of compressor sections | 4 |
| Source vol % $CO_2$ | 99.5 |
| $CO_2$ for transportation (tonnes/year) | 2,007,840 |
| Total compressor electrical power requirement (MWe) | 15.59 |
| Total electrical power requirement of compression cooling water pumps (kWe) | 66 |
| Total Cooling water required (tonnes/hr) | 1949.3 |
| Total Capture plant site area required ($m^2$) | 2600 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 99.6 |
| | $H_2O$ – 0.005 |
| | $N_2$ – 0.14 |
| | $H_2$ – 0.29 |

## Simulation Results

# Scenario 3 – Equipment list

| Summary | | | Equipment Sizing outputs | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Compressor frame | Electrical Duty (MWe) | 15.6 | | | | | 4,445,626 | 72.2 |
| Heat exchanger | Range of heat exchanger heat transfer area ($m^2$) | 12-210 | Number of shell and tube exchangers/double pipe exchangers | 4/1 | Operating pressure ranges bar(a) | 5-41 | 200,405 | 3.3 |
| Knockout drums | Number of knockout drums | 5 | Range of mass of steel required for construction (kg) | 138-8048 | | | 489,561 | 8.0 |
| Electric Drives | Electrical Duty (MWe) | 15.6 | | | | | 230,068 | 3.7 |
| Dehydrator | Water capacity (kg/hr water adsorbed) | 100.52 | | | | | 688,841 | 11.2 |
| Pumps | Electrical Duty (kWe) | 65.6 | | | | | 98,871 | 1.6 |
| Total equipment purchase cost (PCE) | | | | | | | 6,153,372 | |

## Simulation Results
# Scenario 3 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Compressor frame | 4,445,626 | 72.2 |
| Heat exchanger | 200,405 | 3.3 |
| Knockout drums | 489,561 | 8.0 |
| Electric Drives | 230,068 | 3.7 |
| Dehydrator | 688,841 | 11.2 |
| Pumps | 98,871 | 1.6 |
| **Total equipment purchase cost (PCE)** | 6,153,372 | |

# Scenario 3 - Capital Expenditure

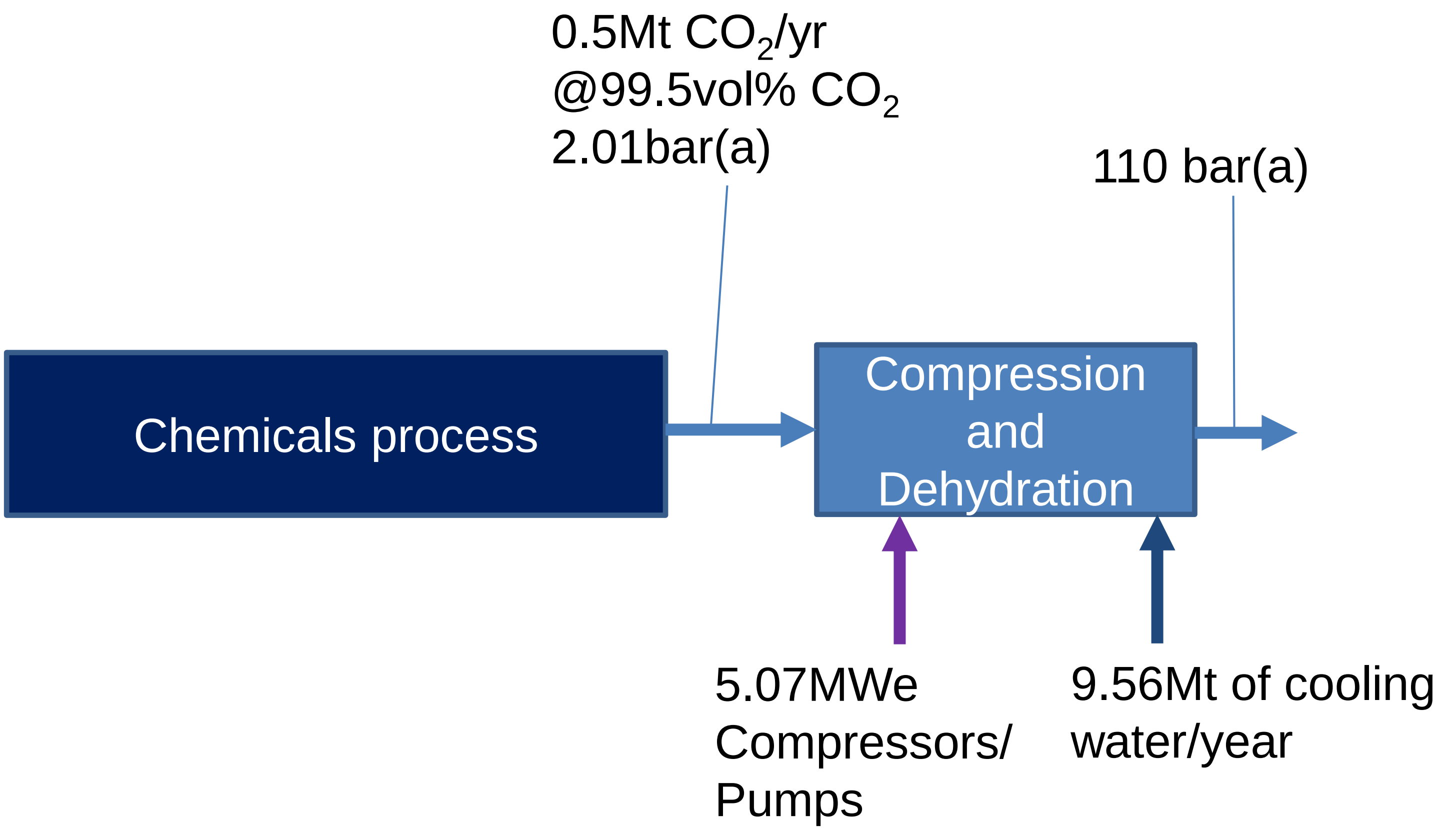
| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 6,153,372 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 19,690,792 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | 38,062,300 |

# Simulation Results
## Scenario 3 - Operating Expenditure

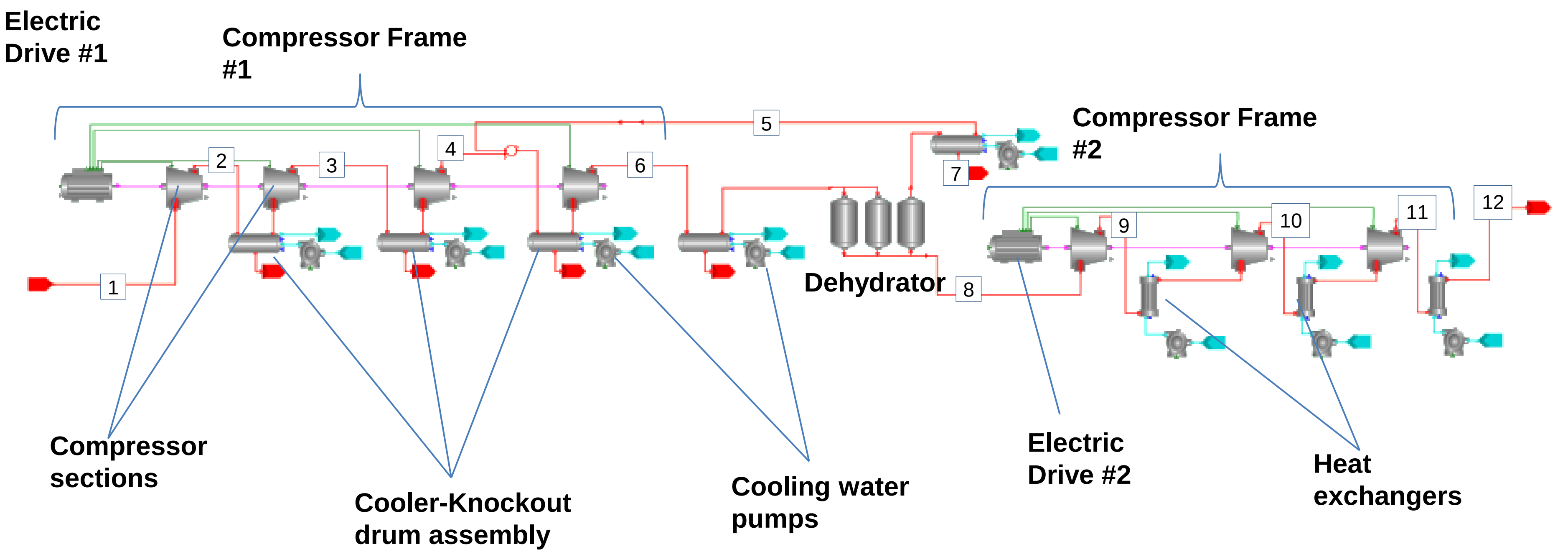
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 1,903,115 |
| Variable Costs | | |
| Miscellaneous materials cost | 10% of maintenance costs | 190,311 |
| Compressor electrical power cost | | 9,913,884 |
| Cooling water pumps power cost | | 41,707 |
| Dehydrator heating costs | | 128,454 |
| Utilities - Cooling water costs | | 408,846 |
| **Total Variable costs** | | 10,683,203 |
| **OPEX** | | 12,586,318 |

## Scenario 4 Process Simulation

# Dense phase, baseline (0.5MT CO2/year)



# Scenario 4 Process Simulation



# Inputs – Scenario 4

- One $CO_2$ compressor train
- 0.5MT $CO_2$/year
- 99.5 vol% $CO_2$ at inlet
- 1bar(g) inlet pressure
- 110bar(a) discharge pressure
- 30°C inlet temperature
- Interstage cooling target – 40°C
- 80Hz compressor frame speed
- Dehydrator moisture specification – 50ppm
- Cooling water temperature is assumed to be 25°C

# Stream Tables
## Scenario 4

| From: | | CO2 SOURCE | FIRST COMPRESSOR | SECOND COMPRESSOR | THIRD COMPRESSOR | RECYCLE COOLERKODRUM | FOURTH COMPRESSOR |
|---|---|---|---|---|---|---|---|
| To: | | FIRST COMPRESSOR | FIRST COOLERKODRUM | SECOND COOLERKODRUM | RECYCLE MIXER | RECYCLE MIXER | FOURTH COOLERKODRUM |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR |
| Stream Number: | | 1 | 2 | 3 | 4 | 5 | 6 |
| **Mass Flow** | **kg/hr** | | | | | | |
| $CO_2$ | | 59,757 | 59,757 | 59,757 | 59,757 | 2,232 | 61,989 |
| $CH_4$ | | 1 | 1 | 1 | 1 | 0 | 1 |
| CO | | 2 | 2 | 2 | 2 | 0 | 2 |
| $H_2O$ | | 25 | 25 | 25 | 25 | 3.5 | 28.1 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 52 | 52 | 52 | 52 | 2 | 54 |
| $H_2$ | | 8 | 8 | 8 | 8 | 0 | 8 |
| $CH_3OH$ | | 1 | 1 | 1 | 1 | 0 | 1 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **TOTAL MASS FLOW** | **kg/hr** | 59,846 | 59,846 | 59,846 | 59,846 | 2,237 | 62,083 |
| Temperature | **ºC** | 30.0 | 119.6 | 116.4 | 100.8 | 40.0 | 93.3 |
| Pressure | **bar(a)** | 2.01 | 5.61 | 12.51 | 23.68 | 23.68 | 41.2 |

# Stream Tables
# Scenario 4

| | | | | | | | |
|---|---|---|---|---|---|---|---|
| From: | | RECYCLE COOLERKODRUM | DEHYDRATOR | FIFTH COMPRESSOR | SIXTH COMPRESSOR | SEVENTH COMPRESSOR | SEVENTH COOLER |
| To: | | WASTE WATER | FIFTH COMPRESSOR | FIFTH COOLERKODRUM | SIXTH COOLERKODRUM | SEVENTH COOLERKODRUM | OUTLET CO2 |
| Service: | | BOTTOMS | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID |
| Phase: | | LIQUID | VAPOUR | VAPOUR | DENSE PHASE | DENSE PHASE | DENSE PHASE |
| Stream Number: | | 7 | 8 | 9 | 10 | 11 | 12 |
| **Mass Flow** | **kg/hr** | | | | | | |
| $CO_2$ | | 0 | 59,757 | 59,757 | 59,757 | 59,757 | 59,757 |
| $CH_4$ | | 0 | 1 | 1 | 1 | 1 | 1 |
| CO | | 0 | 2 | 2 | 2 | 2 | 2 |
| $H_2O$ | | 21.6 | 3 | 3 | 3 | 3 | 3 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 0 | 52 | 52 | 52 | 52 | 52 |
| $H_2$ | | 0 | 8 | 8 | 8 | 8 | 8 |
| $CH_3OH$ | | 0 | 1 | 1 | 1 | 1 | 1 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **TOTAL MASS FLOW** | **kg/hr** | 22 | 59,824 | 59,824 | 59,824 | 59,824 | 59,824 |
| Temperature | **ºC** | 40.0 | 40.3 | 96.6 | 67.6 | 44.3 | 40.0 |
| Pressure | **bar(a)** | 23.7 | 40.0 | 71.5 | 98.1 | 110.2 | 110.0 |

## Simulation Results

# Scenario 4 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of compression | 1 |
| Number of compressor frames | 2 |
| Number of compressor sections | 7 |
| Source vol % $CO_2$ | 99.5 |
| $CO_2$ for transportation (tonnes/year) | 501,961 |
| Total compressor electrical power requirement (MWe) | 5.03 |
| Total electrical power requirement of compression cooling water pumps (kWe) | 38 |
| Total Cooling water required (tonnes/hr) | 1138.5 |
| Total Capture plant site area required ($m^2$) | 4550 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 99.6 |
| | $H_2O$ – 0.005 |
| | $N_2$ – 0.14 |
| | $H_2$ – 0.29 |

# Simulation Results

## Scenario 4 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Compressor frame | Electrical Duty (MWe) | 5.0 | | | | | 3,329,105 | 50.5 |
| Heat exchanger | Range of heat exchanger heat transfer area ($m^2$) | 3-116 | Number of shell and tube exchangers/double pipe exchangers | 7/1 | Operating pressure ranges bar(a) | 5-110 | 267,173 | 26.6 |
| Knockout drums | Number of knockout drums | 5 | Range of mass of steel required for construction (kg) | 120-1418 | | | 146,991 | 6.2 |
| Electric Drives | Electrical Duty (MWe) | 5.0 | | | | | 177,518 | 3.4 |
| Dehydrator | Water capacity (kg/hr water adsorbed) | 25.13 | | | | | 322,006 | 2.5 |
| Pumps | Electrical Duty (kWe) | 38.3 | | | | | 73,272 | 1.7 |
| Total equipment purchase cost (PCE) | | | | | | | 4,316,066 | |

# Scenario 4 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Compressor frame | 3,329,105 | 50.5 |
| Heat exchanger | 267,173 | 26.6 |
| Knockout drums | 146,991 | 6.2 |
| Electric Drives | 177,518 | 3.4 |
| Dehydrator | 322,006 | 2.5 |
| Pumps | 73,272 | 1.7 |
| **Total equipment purchase cost (PCE)** | 4,316,066 | |

# Scenario 4 - Capital Expenditure

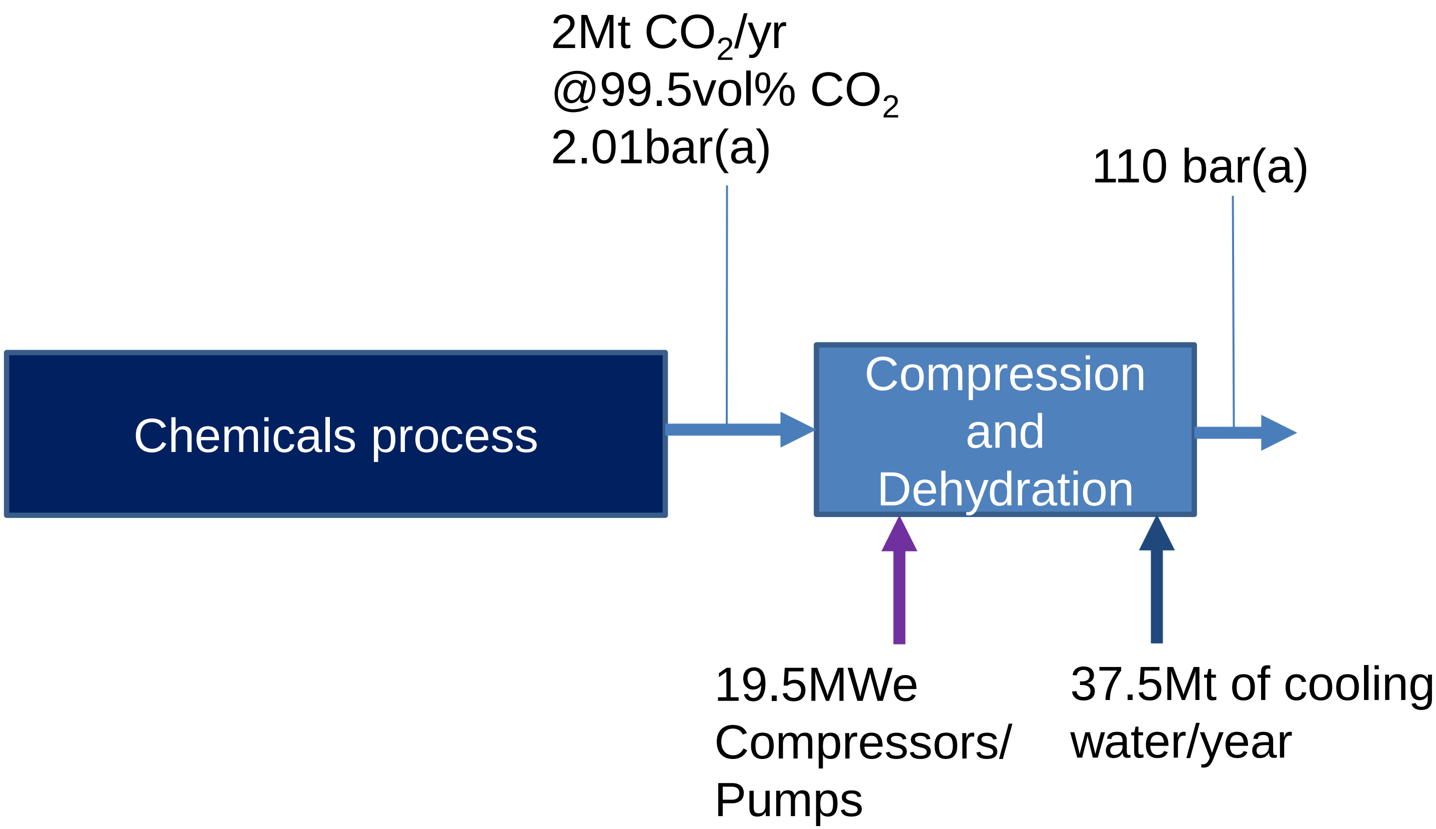
| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 4,316,066 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 13,811,410 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | 26,697,455 |

# Scenario 4 - Operating Expenditure

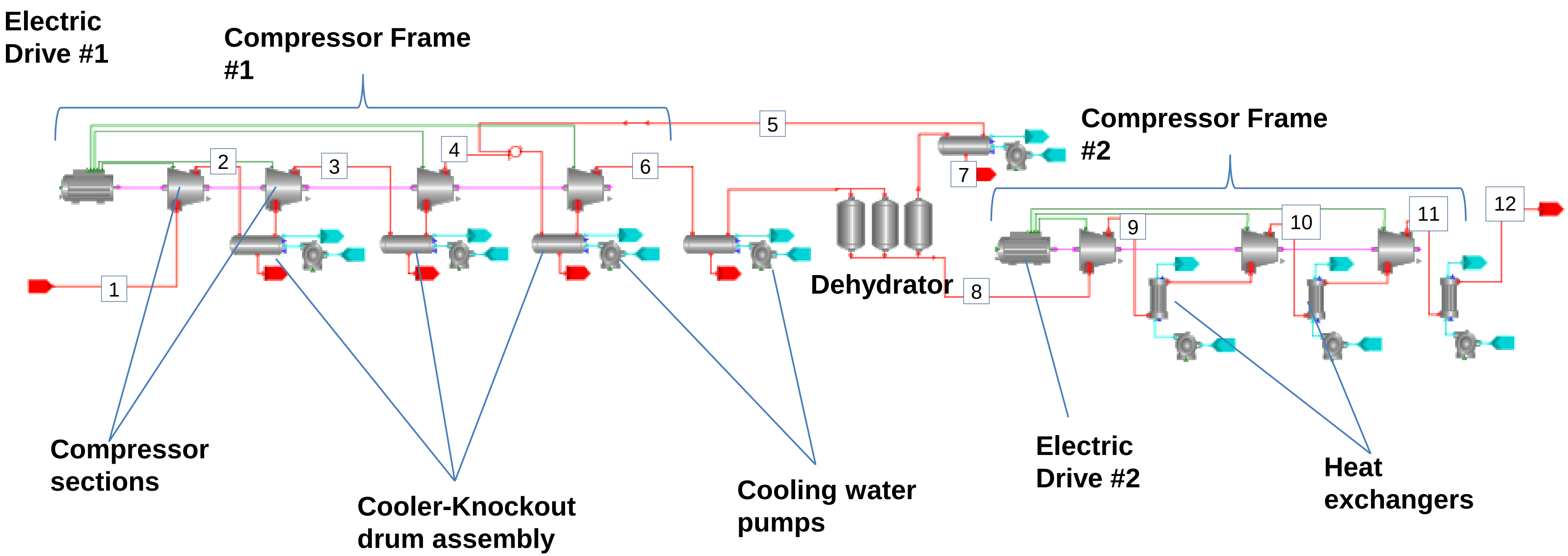
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 1,334,873 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 133,487 |
| Compressor electrical power cost | | 3,093,281 |
| Cooling water pumps power cost | | 23,554 |
| Dehydrator heating costs | | 31,933 |
| Utilities - Cooling water costs | | 218,931 |
| **Total Variable costs** | | 3,501,187 |
| **OPEX** | | 4,836,060 |

## Scenario 5 Process Simulation

# Dense phase, large (2Mt $CO_2$/year)



# Scenario 5 Process Simulation



# Inputs – Scenario 5

- One $CO_2$ compressor train
- 2MT $CO_2$/year
- 99.5 vol% $CO_2$ at inlet
- 1bar(g) inlet pressure
- 110bar(a) discharge pressure
- 30°C inlet temperature
- Interstage cooling target – 40°C
- 80Hz compressor frame speed
- Dehydrator moisture specification – 50ppm
- Cooling water temperature is assumed to be 25°C

# Scenario 5

| From: | | CO2 SOURCE | FIRST COMPRESSOR | SECOND COMPRESSOR | THIRD COMPRESSOR | RECYCLE COOLERKODRUM | FOURTH COMPRESSOR |
|---|---|---|---|---|---|---|---|
| To: | | FIRST COMPRESSOR | FIRST COOLERKODRUM | SECOND COOLERKODRUM | RECYCLE MIXER | RECYCLE MIXER | FOURTH COOLERKODRUM |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | LIQUID | VAPOUR | VAPOUR | VAPOUR |
| Stream Number: | | 1 | 2 | 3 | 4 | 5 | 6 |
| **Mass Flow** | kg/hr | | | | | | |
| $CO_2$ | | 239,029 | 239,029 | 239,029 | 239,029 | 8,926 | 247,955 |
| $CH_4$ | | 4 | 4 | 4 | 4 | 0 | 5 |
| CO | | 8 | 8 | 8 | 8 | 0 | 8 |
| $H_2O$ | | 98 | 98 | 98 | 98 | 14.1 | 112.5 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 206 | 206 | 206 | 206 | 8 | 214 |
| $H_2$ | | 32 | 32 | 32 | 32 | 1 | 33 |
| $CH_3OH$ | | 4 | 4 | 4 | 4 | 0 | 4 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 239,382 | 239,382 | 239,382 | 239,382 | 8,950 | 248,332 |
| Temperature | ºC | 30.0 | 117.2 | 113.7 | 98.9 | 40.0 | 92.0 |
| Pressure | bar(a) | 2.01 | 5.61 | 12.51 | 23.68 | 23.68 | 41.2 |

# Stream Tables
# Scenario 5

| From: | | RECYCLE COOLERKODRUM | DEHYDRATOR | FIFTH COMPRESSOR | SIXTH COMPRESSOR | SEVENTH COMPRESSOR | SEVENTH COOLER |
|---|---|---|---|---|---|---|---|
| To: | | WASTE WATER | FIFTH COMPRESSOR | FIFTH COOLERKODRUM | SIXTH COOLERKODRUM | SEVENTH COOLERKODRUM | OUTLET CO2 |
| Service: | | BOTTOMS | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID |
| Phase: | | LIQUID | VAPOUR | VAPOUR | DENSE PHASE | DENSE PHASE | DENSE PHASE |
| Stream Number: | | 7 | 8 | 9 | 10 | 11 | 12 |
| **Mass Flow** | kg/hr | | | | | | |
| $CO_2$ | | 0 | 239,029 | 239,029 | 239,029 | 239,029 | 239,029 |
| $CH_4$ | | 0 | 4 | 4 | 4 | 4 | 4 |
| CO | | 0 | 8 | 8 | 8 | 8 | 8 |
| $H_2O$ | | 86.4 | 12 | 12 | 12 | 12 | 12 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 0 | 206 | 206 | 206 | 206 | 206 |
| $H_2$ | | 0 | 32 | 32 | 32 | 32 | 32 |
| $CH_3OH$ | | 0 | 4 | 4 | 4 | 4 | 4 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 87 | 239,296 | 239,296 | 239,296 | 239,296 | 239,296 |
| Temperature | ºC | 40.0 | 40.3 | 95.5 | 67.0 | 44.3 | 40.0 |
| Pressure | bar(a) | 23.7 | 40.0 | 71.5 | 98.1 | 110.2 | 110.0 |

# Scenario 5 - Process Conditions

| Description | Value |
| --- | --- |
| Number of trains of compression | 1 |
| Number of compressor frames | 2 |
| Number of compressor sections | 7 |
| Source vol % $CO_2$ | 99.5 |
| $CO_2$ for transportation (tonnes/year) | 2,007,840 |
| Total compressor electrical power requirement (MWe) | 19.35 |
| Total electrical power requirement of compression cooling water pumps (kWe) | 150 |
| Total Cooling water required (tonnes/hr) | 4464.0 |
| Total Capture plant site area required ($m^2$) | 4550 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 99.6 |
| | $H_2O$ – 0.005 |
| | $N_2$ – 0.14 |
| | $H_2$ – 0.29 |

# Simulation Results

# Scenario 5 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Compressor frame | Electrical Duty (MWe) | 19.4 | | | | | 6,551,213 | 49.5 |
| Heat exchanger | Range of heat exchanger heat transfer area ($m^2$) | 12-467 | Number of shell and tube exchangers/double pipe exchangers | 7/1 | Operating pressure ranges bar(a) | 5-110 | 619,785 | 23.4 |
| Knockout drums | Number of knockout drums | 5 | Range of mass of steel required for construction (kg) | 138-8048 | | | 489,561 | 6.9 |
| Electric Drives | Electrical Duty (MWe) | 19.4 | | | | | 334,844 | 5.4 |
| Dehydrator | Water capacity (kg/hr water adsorbed) | 100.52 | | | | | 688,841 | 2.6 |
| Pumps | Electrical Duty (kWe) | 150.2 | | | | | 302,883 | 1.2 |
| Total equipment purchase cost (PCE) | | | | | | | 8,987,128 | |

# Scenario 5 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Compressor frame | 6,551,213 | 49.5 |
| Heat exchanger | 619,785 | 23.4 |
| Knockout drums | 489,561 | 6.9 |
| Electric Drives | 334,844 | 5.4 |
| Dehydrator | 688,841 | 2.6 |
| Pumps | 302,883 | 1.2 |
| **Total equipment purchase cost (PCE)** | 8,987,128 | |

# Scenario 5 - Capital Expenditure

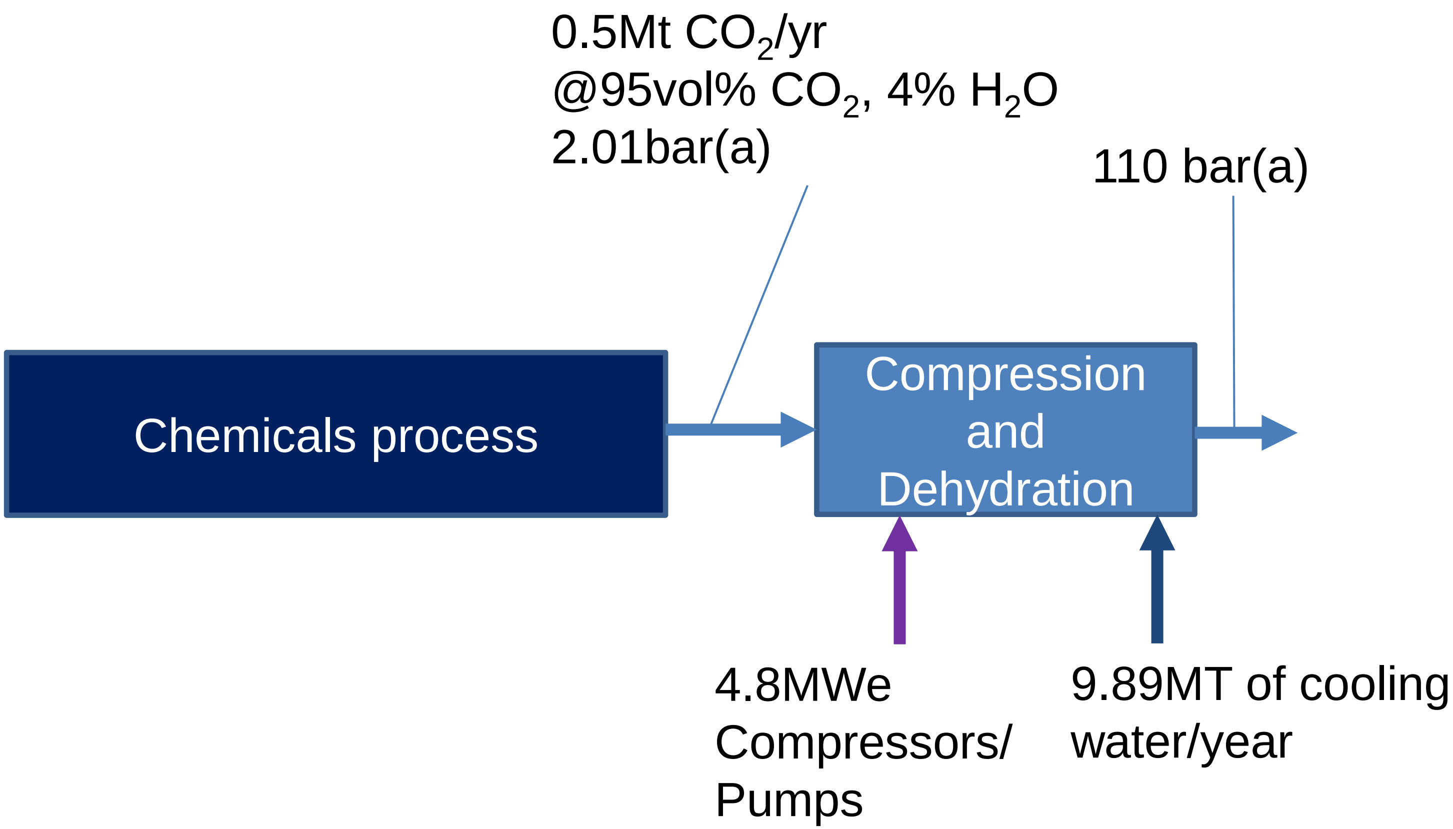
| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 8,987,128 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 28,758,809 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | 55,590,777 |

# Scenario 5 - Operating Expenditure

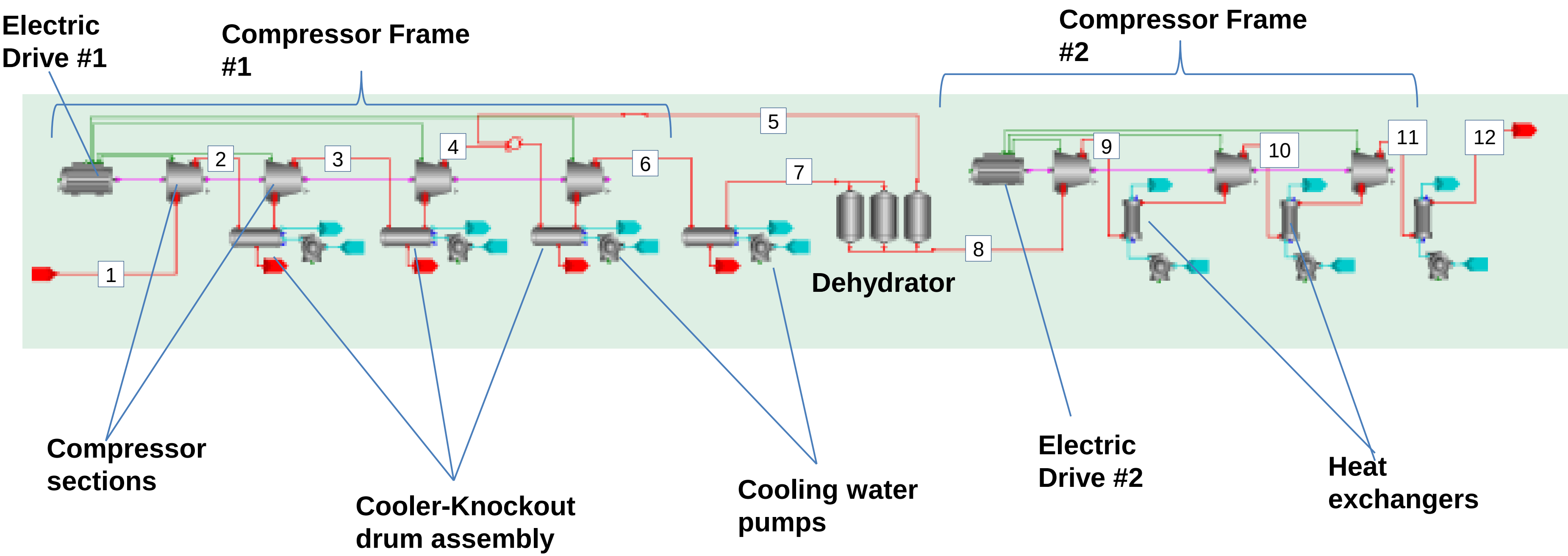
| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 2,779,539 |
| **Variable Costs** | | |
| Miscellaneous materials cost | 10% of maintenance costs | 277,954 |
| Compressor electrical power cost | | 12,306,994 |
| Cooling water pumps power cost | | 95,513 |
| Dehydrator heating costs | | 128,454 |
| Utilities - Cooling water costs | | 858,448 |
| **Total Variable costs** | | 13,667,362 |
| **OPEX** | | 16,446,901 |

## Scenario 6 Process Simulation

# Dense phase, low purity $CO_2$ (0.5Mt $CO_2$/year)



# Scenario 6 Process Simulation



# Inputs – Scenario 6

- One $CO_2$ compressor train
- 0.5MT $CO_2$/year
- 95 vol% $CO_2$ at inlet
- 1bar(g) inlet pressure
- 110bar(a) discharge pressure
- 30°C inlet temperature
- Interstage cooling target – 40°C
- 80Hz compressor frame speed
- Dehydrator moisture specification – 50ppm
- Cooling water temperature is assumed to be 25°C

# Stream Tables
## Scenario 6

| From: | | CO2 SOURCE | FIRST COMPRESSOR | SECOND COMPRESSOR | THIRD COMPRESSOR | RECYCLE COOLERKODRUM | FOURTH COMPRESSOR |
|---|---|---|---|---|---|---|---|
| To: | | FIRST COMPRESSOR | FIRST COOLERKODRUM | SECOND COOLERKODRUM | RECYCLE MIXER | RECYCLE MIXER | FOURTH COOLERKODRUM |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR | VAPOUR |
| Stream Number: | | 1 | 2 | 3 | 4 | 5 | 6 |
| **Mass Flow** | kg/hr | | | | | | |
| $CO_2$ | | 61,192 | 61,192 | 61,192 | 61,192 | 2,285 | 63,477 |
| $CH_4$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| CO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $H_2O$ | | 1,055 | 1,055 | 363 | 169 | 66.9 | 101.8 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 410 | 410 | 410 | 410 | 15 | 425 |
| $H_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $CH_3OH$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 62,657 | 62,657 | 61,965 | 61,770 | 2,367 | 64,004 |
| Temperature | ºC | 30.0 | 90.2 | 116.5 | 100.9 | 200.8 | 93.4 |
| Pressure | bar(a) | 2.01 | 5.61 | 12.51 | 23.68 | 23.68 | 41.2 |

# Stream Tables
## Scenario 6

| From: | | FOURTH COOLERKODRUM | DEHYDRATOR | FIFTH COMPRESSOR | SIXTH COMPRESSOR | SEVENTH COMPRESSOR | SEVENTH COOLER |
|---|---|---|---|---|---|---|---|
| To: | | DEHYDRATOR | FIFTH COMPRESSOR | FIFTH COOLER | SIXTH COOLER | SEVENTH COOLER | OUTLET CO2 |
| Service: | | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID | CO2 FLUID |
| Phase: | | VAPOUR | VAPOUR | VAPOUR | DENSE PHASE | DENSE PHASE | DENSE PHASE |
| Stream Number: | | 7 | 8 | 9 | 10 | 11 | 12 |
| **Mass Flow** | kg/hr | | | | | | |
| $CO_2$ | | 63,477 | 61,192 | 61,192 | 61,192 | 61,192 | 61,192 |
| $CH_4$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| CO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $H_2O$ | | 70.0 | 3 | 3 | 3 | 3 | 3 |
| $H_2S$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| NO | | 0 | 0 | 0 | 0 | 0 | 0 |
| $O_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $N_2$ | | 425 | 410 | 410 | 410 | 410 | 410 |
| $H_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $CH_3OH$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| $SO_2$ | | 0 | 0 | 0 | 0 | 0 | 0 |
| Particulates | | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **TOTAL MASS FLOW** | kg/hr | 63,972 | 61,605 | 61,605 | 61,605 | 61,605 | 61,605 |
| Temperature | ºC | 40.0 | 42.5 | 99.1 | 67.7 | 44.6 | 40.0 |
| Pressure | bar(a) | 41.0 | 40.0 | 71.5 | 98.1 | 110.2 | 110.0 |

## Simulation Results

# Scenario 6 - Process Conditions

| Description | Value |
|---|---|
| Number of trains of compression | 1 |
| Number of compressor frames | 2 |
| Number of compressor sections | 7 |
| Source vol % $CO_2$ | 95.0 |
| $CO_2$ for transportation (tonnes/year) | 514,012 |
| Total compressor electrical power requirement (MWe) | 4.76 |
| Total electrical power requirement of compression cooling water pumps (kWe) | 40 |
| Total Cooling water required (tonnes/hr) | 1177.2 |
| Total Capture plant site area required ($m^2$) | 4550 |
| Output $CO_2$ stream conditions (vol%) | $CO_2$ – 99.0 |
| | $H_2O$ – 0.005 |
| | $N_2$ – 1.0 |
| | $H_2$ – 0.0 |

## Simulation Results

# Scenario 6 – Equipment list

| Summary | Equipment Sizing outputs | | | | | | £ | % |
|---|---|---|---|---|---|---|---|---|
| Compressor frame | Electrical Duty (MWe) | 4.8 | | | | | 3,263,257 | 47.5 |
| Heat exchanger | Range of heat exchanger heat transfer area ($m^2$) | 36-114 | Number of shell and tube exchangers/double pipe exchangers | 7/0 | Operating pressure ranges bar(a) | 5-110 | 238,583 | 26.5 |
| Knockout drums | Number of knockout drums | 4 | Range of mass of steel required for construction (kg) | 1322-1447 | | | 136,376 | 5.4 |
| Electric Drives | Electrical Duty (MWe) | 4.8 | | | | | 177,326 | 3.1 |
| Dehydrator | Water capacity (kg/hr water adsorbed) | 66.95 | | | | | 523,295 | 2.4 |
| Pumps | Electrical Duty (kWe) | 39.6 | | | | | 69,024 | 1.7 |
| Total equipment purchase cost (PCE) | | | | | | | 4,407,861 | |

# Scenario 6 - Capital Expenditure

| Description | £ | % of PCE |
|---|---|---|
| **Equipment purchase cost breakdown** | | |
| Compressor frame | 3,263,257 | 47.5 |
| Heat exchanger | 238,583 | 26.5 |
| Knockout drums | 136,376 | 5.4 |
| Electric Drives | 177,326 | 3.1 |
| Dehydrator | 523,295 | 2.4 |
| Pumps | 69,024 | 1.7 |
| **Total equipment purchase cost (PCE)** | 4,407,861 | |

# Scenario 6 - Capital Expenditure

| Description | Factor (%) | Cost (£) (Q3 2013) |
|---|---|---|
| **Total purchase cost (PCE)** | | 4,407,861 |
| **Supply of materials** | | |
| Foundations and paving | 10 | |
| Platforms and supporting | 15 | |
| Buildings | | |
| Piping | 60 | |
| Insulation and fireproofing | 25 | |
| Electrical | 5 | |
| Painting cleaning | | |
| Testing and miscellaneous | 3 | |
| **Transport and installation** | | |
| Transport and installation of equipment | 10 | |
| Installation of materials | 72 | |
| US prices to European | 20 | |
| **Total Plant installed capital cost** | | 14,105,156 |
| Contingency | 30 | |
| Design and engineering | 30 | |
| Indirect cost (project management, permitting, taxes) | 33.3 | |
| **Total fixed capital cost** | | 27,265,267 |

# Scenario 6 - Operating Expenditure

| Description | | £/year |
|---|---|---|
| **Fixed Costs** | | |
| Maintenance, Staff, Insurance and Overheads | 5% of Fixed Capital | 1,363,263 |
| Variable Costs | | |
| Miscellaneous materials cost | 10% of maintenance costs | 136,326 |
| Compressor electrical power cost | | 2,926,391 |
| Cooling water pumps power cost | | 24,356 |
| Dehydrator heating costs | | 32,542 |
| Utilities - Cooling water costs | | 226,384 |
| **Total Variable costs** | | 3,345,999 |
| **OPEX** | | 4,709,263 |

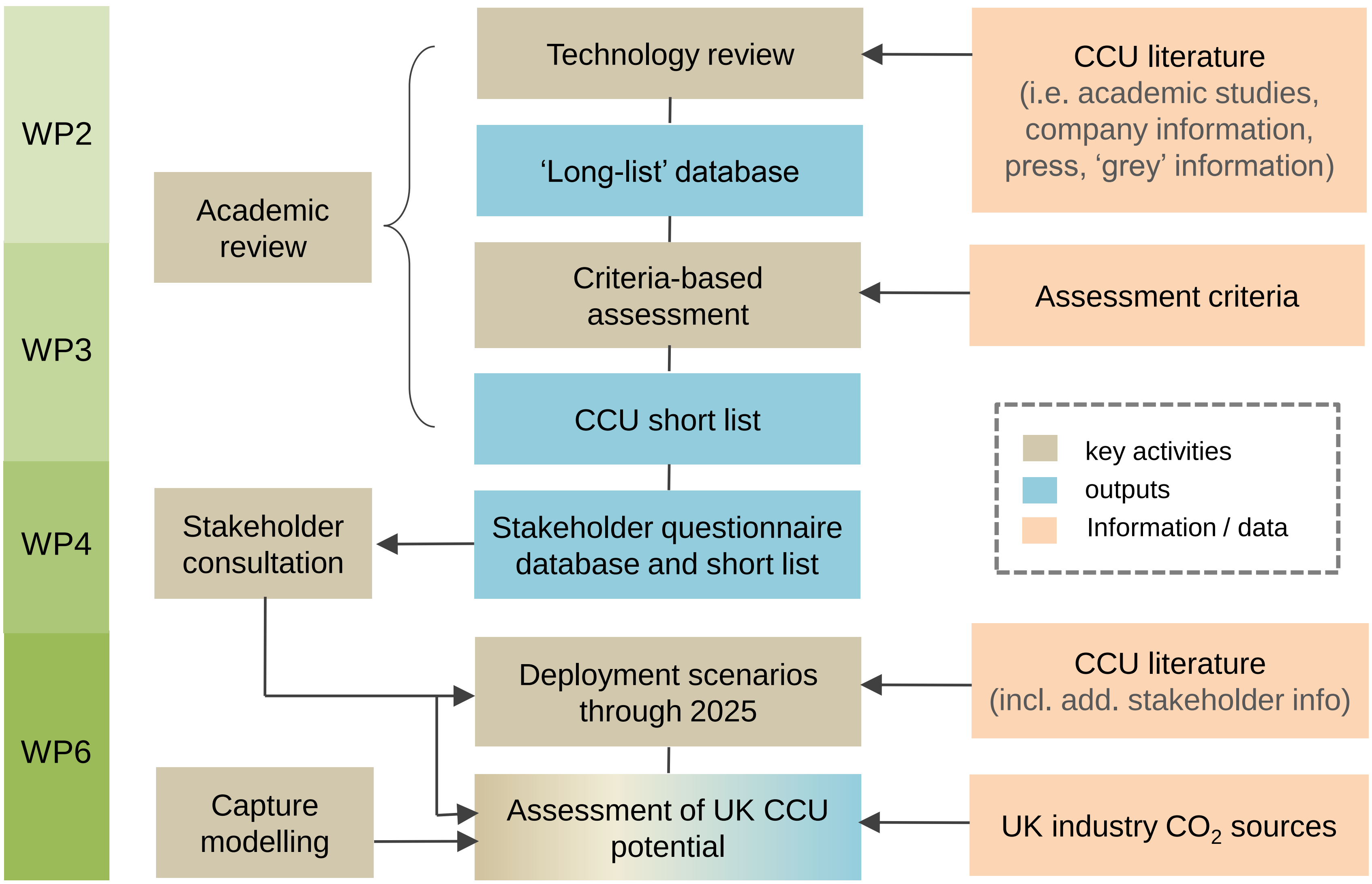
# Comparison of costs between scenarios

| | Baseline Scenario #1 | Sensitivity "Gas phase, Pilot" (Scenario #2) | Sensitivity "Gas phase, Large scale" (Scenario #3) | Sensitivity "Dense phase, Baseline" (Scenario #4) | Sensitivity "Dense phase, large scale" (Scenario #5) | Sensitivity "Dense phase, low purity $CO_2$" (Scenario #6) |
|---|---|---|---|---|---|---|
| Source $CO_2$ | 0.5 Mt/yr | 0.1 Mt/yr | 2 Mt/yr | 0.5 Mt/yr | 2 Mt/yr | 0.5 Mt/yr |
| Equipment cost | £3m | £1.5m | £6m | £4m | £9m | £4m |
| Total fixed cost | £15m | £8m | £32m | £22m | £47m | £23m |
| Annual opex (incl. energy) | £5m/yr | £1m/yr | £13m/yr | £5m/yr | £16m/yr | £5m/yr |
| Power /$MW_e$ | 4.2 | 0.89 | 16.4 | 9.6 | 37.5 | 9.9 |

# Outline

- Overall Project Methodology
- $CO_2$ capture technologies
- $CO_2$ sources
- Techno-economic analysis of industrial $CO_2$ capture
- Process simulation case studies
- $CO_2$ utilisation review

# Approach to the $CO_2$ utilisation work-stream



# Sources of information and data limitations

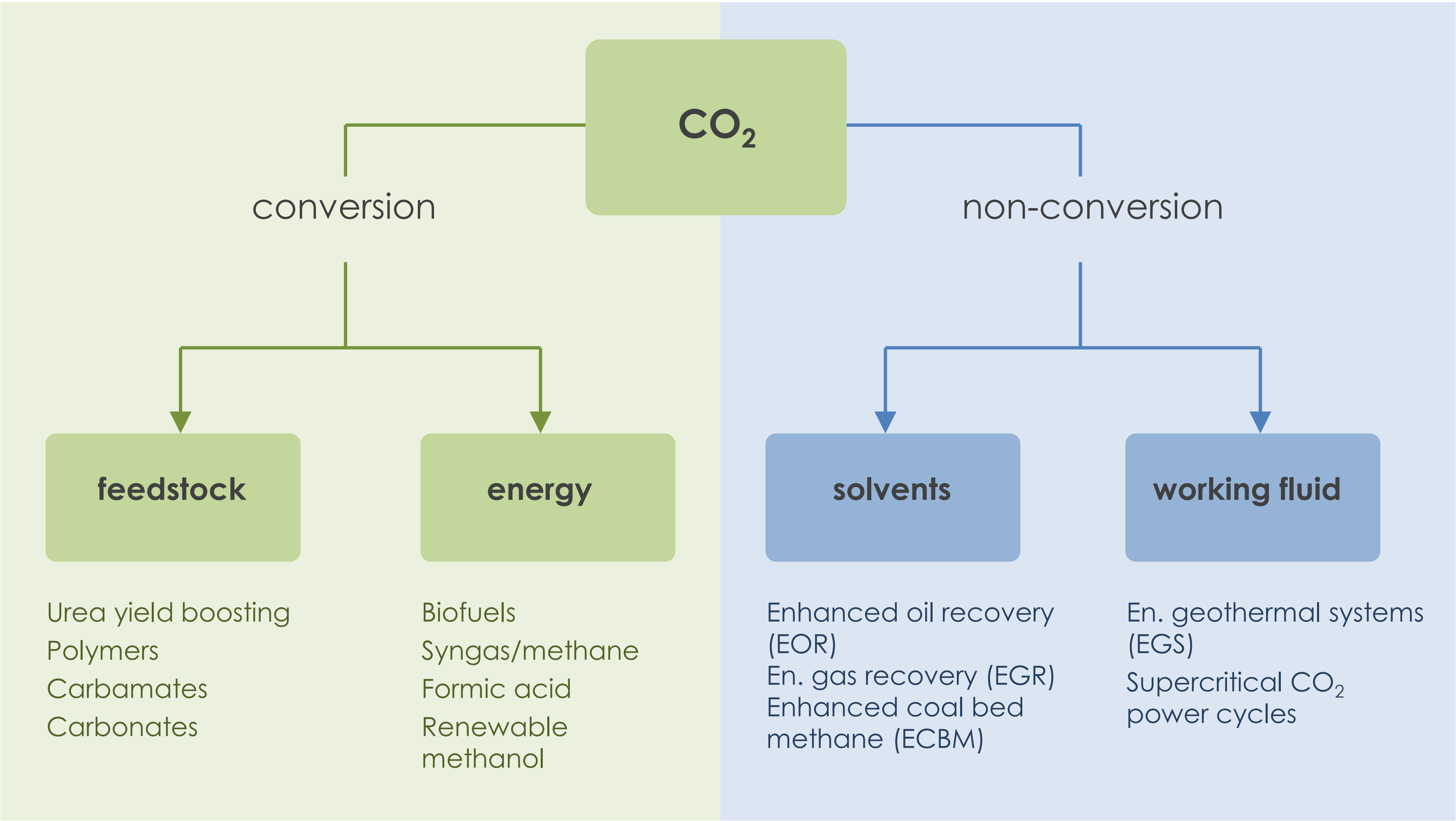
WP2 technology review based on latest information on CCU technology:

- Recent global studies on CCU (e.g. Carbon Sequestration Leadership Forum (CSLF), 2011. Phase I Final Report by the CSLF Task Force on $CO_2$ Utilization; CSLF, 2013 Phase II Final Report; and Global Carbon Capture and Storage Institute (GCCSI), 2011. Accelerating the uptake of CCS: Industrial use of captured $CO_2$)
- Academic literature (extensive body; largely based around early R&D activities)
- Company information (e.g. project and process info from start-ups and multinationals)
- Press and trade assoc. (project info within trade and specialised press, and various trade groups)
- CCU technology networks and activities (e.g. CO2chem; International Conference on Carbon Dioxide Utilization (ICCDU); Foreseeing a future using $CO_2$ (4CU); Supercritical $CO_2$ Power Cycle Symposium (SCO2PCS))

Data limitations and challenges:

- Majority of CCU technologies are at early R&D stages (TRL 1-3); much technical information but limited to small-scale lab tests - with little or no economic data
- For more mature uses of $CO_2$ (other than EOR), almost all cost data is confidential
- Performance and (limited) cost data published by companies typically not supported by key assumptions, boundary systems etc; optimistic claims also need to be viewed with some caution and objectivity!
- Recently started projects should improve the dataset (for some CCU technologies) - but have yet to report (e.g. EC-JRC study, Smart $CO_2$ transformation (SCOT)
- Stakeholder consultation aims to uncover any further sources of relevant data

# Context: What is CCU?




Project considers both $CO_2$ utilisation (involving conversion) and $CO_2$ uses (excl. EOR)

# Stakeholder views as to why is CCU of potential interest to the UK

## Drivers for considering CCU in the UK:

- Support UK industrial innovation and competitiveness
  - Action #2 of the Government's approach to Industrial Strategy[1] is to *support emerging technologies*
  - Report on UK competitiveness report 'No Stone Unturned' support for new tech.
  - CCU identified as one of the Top 10 emerging technology trends by the WEF[2]
- Emergence of new techniques to convert $CO_2$ to high value products (i.e. use of waste material for commercial production)
- Ability to enhance energy security and support renewable energy (including energy storage/link to UK offshore wind strategy etc)
- Concerns over CCS value-chain costs/lack of progress (CCU as a support to CCS)

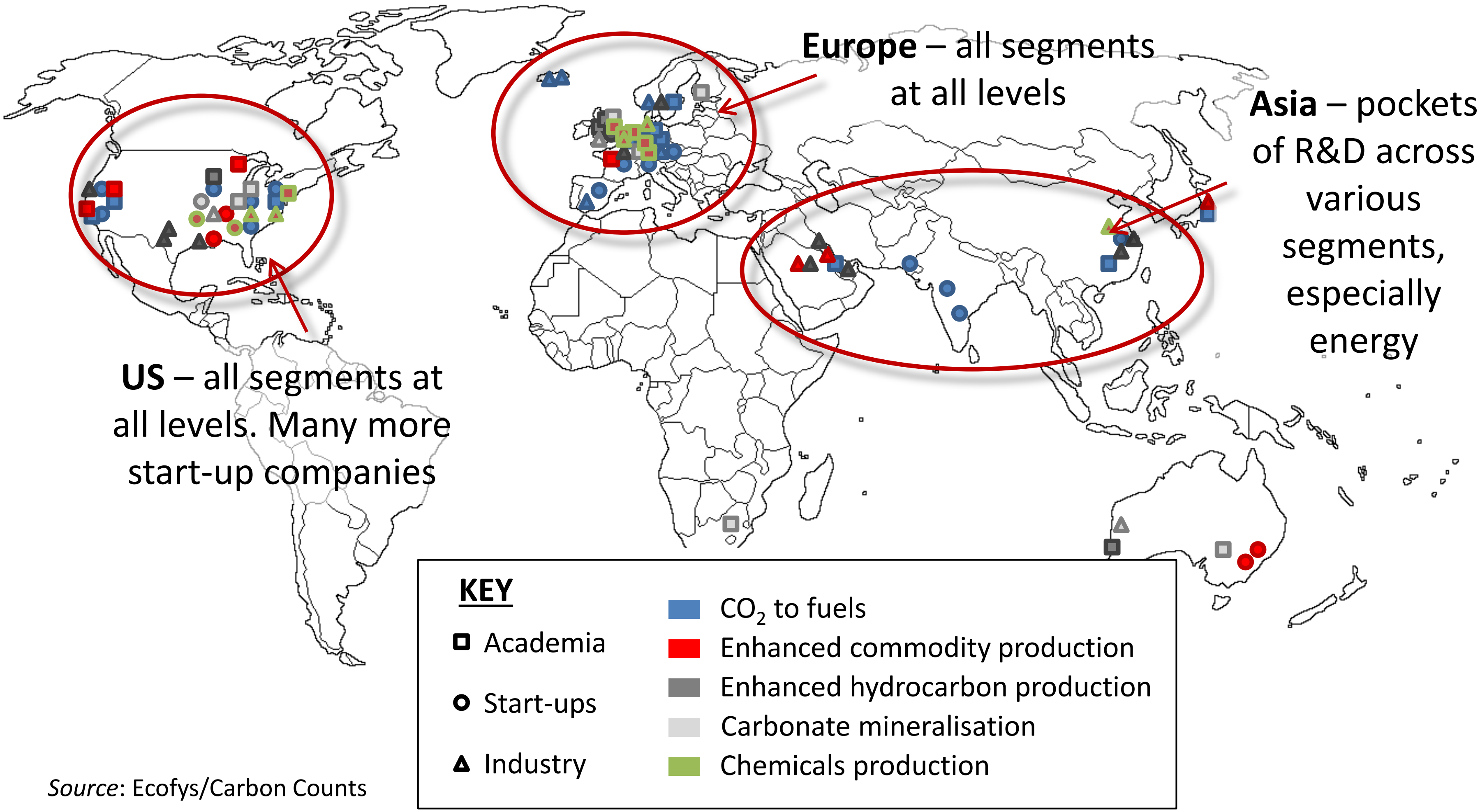
## Challenges and barriers:

- Low activation state, therefore need for energy + catalysts = costs, additional energy and emissions; various other barriers depending upon product/sector/market
- Focus of R&D across most pre-commercial CCU applications is therefore around increasing process efficiency and energy optimisation; also need for scale-up to demonstration technology improve economics

1. 'Using Industrial Strategy to help the UK economy and business compete and grow' (BIS, Sept 2012)
2. Global Agenda Council on Emerging Technologies 2012-2014, World Economic Forum

# Who is involved in CCU development?

- Geographical factors play an important role in determining interest/potential for many CCU technologies (e.g. climate, material/energy availability etc.)




*Source*: Ecofys/Carbon Counts

# What is the current status of CCU technology?

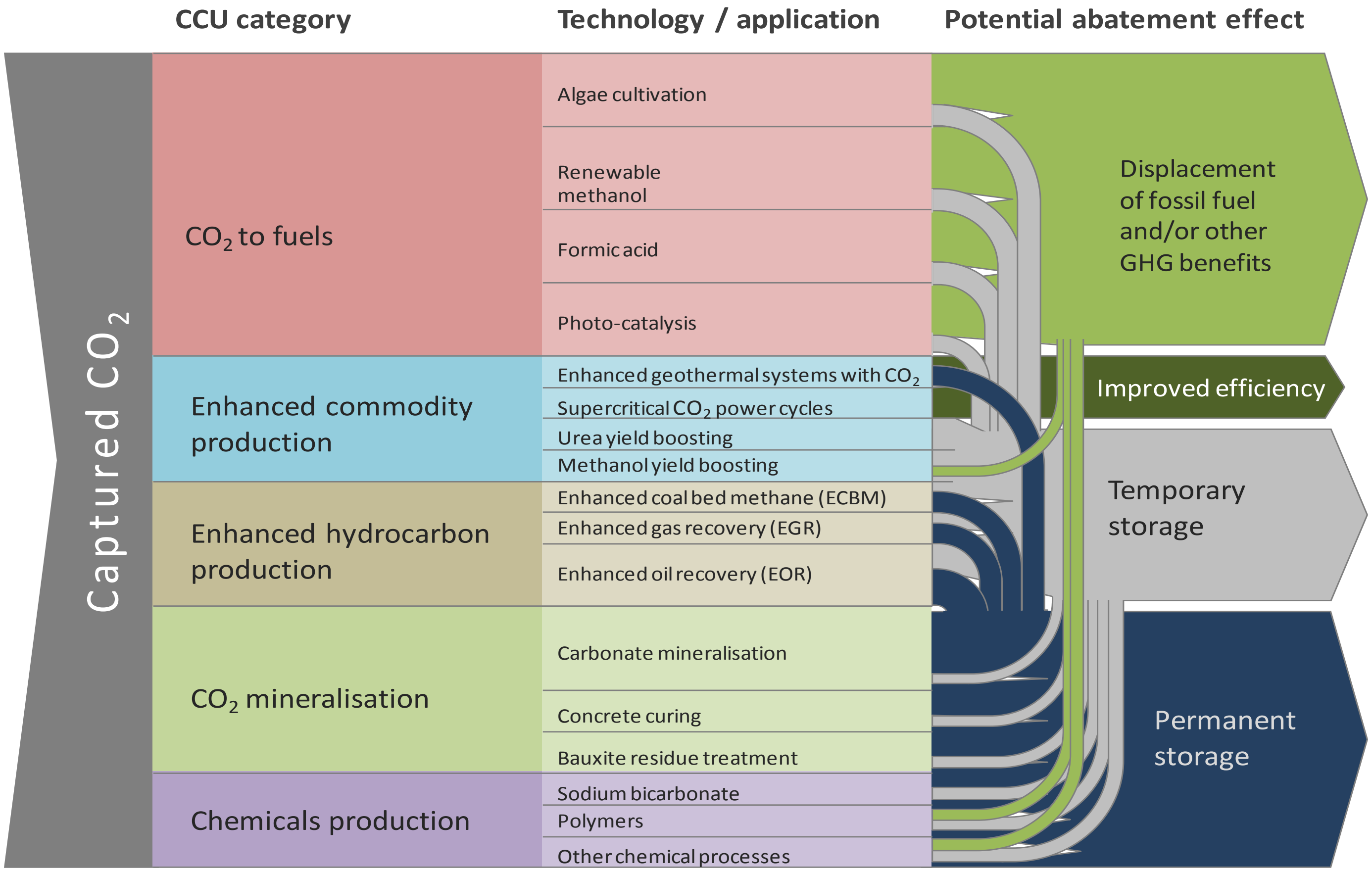
| CCU category | CCU technology | Research | Demon-stration | Economically feasible under certain conditions | Mature market |
|---|---|---|---|---|---|
| $CO_2$ to fuels | Hydrogen (renewable methanol) | | Main activities | Some activities | |
| | Hydrogen (formic acid) | Main activities | Some activities | | |
| | Algae (to biofuels) | Main activities | Main activities | | |
| | Photocatalytic processes | Main activities | | | |
| | Nanomaterial catalysts | Main activities | | | |
| Enhanced commodity production | Power cycles (using $scCO_2$) | Main activities | | | |
| | Enhanced production (urea; methanol) | | | | Main activities |
| Enhanced hydrocarbon recovery | Miscible/immiscible floods ($CO_2$-EOR) | | | Some activities | Main activities |
| | Miscible/immiscible floods ($CO_2$-EGR) | | Main activities | | |
| | Sorption-based displacement (ECBM) | Main activities | Some activities | | |
| $CO_2$ mineralisation | Carbonate mineralisation | | Some activities | Main activities | |
| | $CO_2$ concrete curing | | | Main activities | |
| | Bauxite residue carbonation | | | Main activities | |
| Chemicals production | Sodium carbonate | | | | Main activities |
| | Polymers | Main activities | Some activities | | |
| | Other chemicals (e.g. acetic acid) | Main activities | | | |
| | Algae (for chemicals) | | Main activities | Some activities | |

*Main activities*
*Some activities*

**Urea and EOR accounts for almost all $CO_2$ use to date globally**

*Source*: Ecofys/Carbon Counts

# CCU and climate policy: various routes to GHG reduction

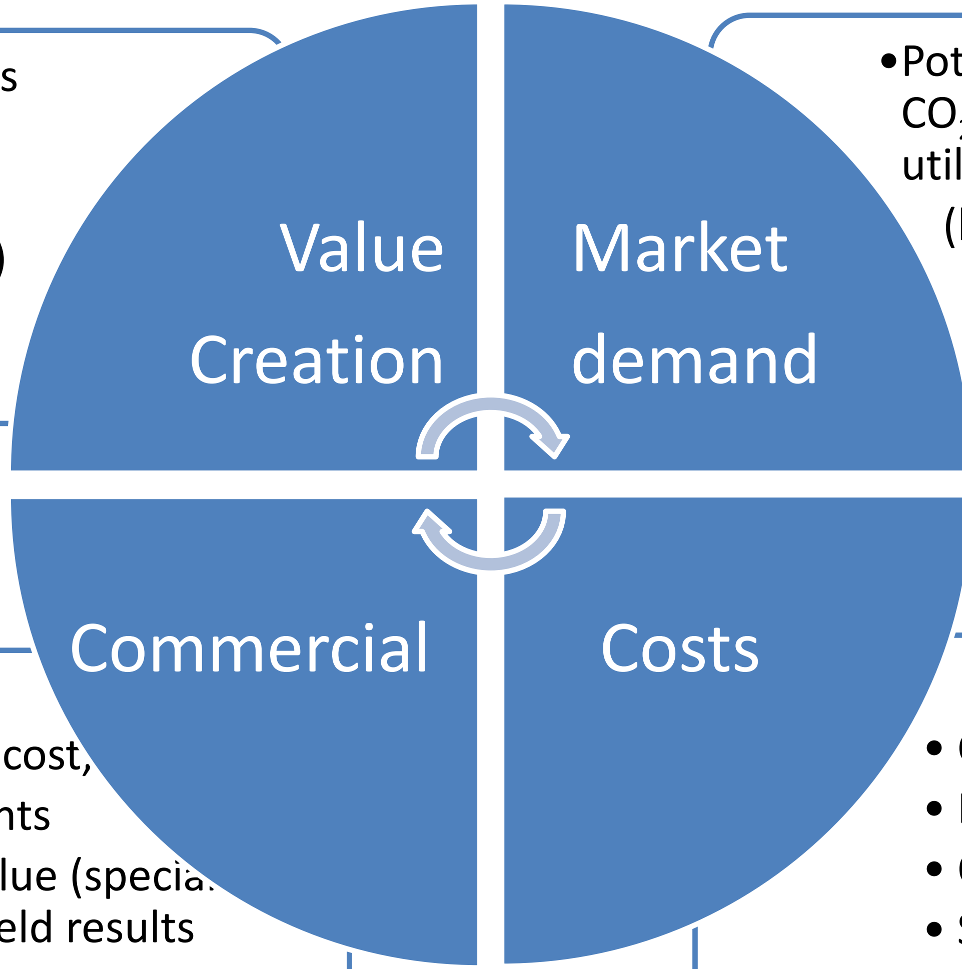



*Source*: Ecofys/Carbon Counts

- Wide range of potential $CO_2$ abatement effects across options + project settings
- GHG LCA impacts dependent upon range of factors e.g. energy source; products etc
- MRGs no longer allow transfers to be deducted from inventory; however, scope exists to include *"...future innovations"* to allow new pathways for CCU to be opted-in to EU ETS

# Economic factors affecting CCU

- Creating revenues (using waste)
- Avoiding costs (e.g. EU ETS costs)




- Potential volumes of $CO_2$ that could be utilised (highly variable <1 $MtCO_2$/yr to several 100)

- Displacing lower cost, trusted, incumbents
- Focus on high-value (specia products could yield results
- Eco-labelling

- Capital
- Energy
- Other OPEX
- Scope for cost reduction

# CCU assessment criteria

Which CCU options could be of greatest applicability to the UK through 2025?

Traffic-light assessment of ‘long-list’ based on 3 key areas:

A. Technology development and performance
   - Technology readiness level (TRL)
   - Energy performance (including energy storage potential)
   - Abatement potential (e.g. permanent vs. temp. storage; fossil fuel substitution)
   - Environmental, heath and safety factors/concerns (non-GHG)

B. Economic and commercial potential
   - Uptake potential (size of potential market)
   - Economic potential (various cost factors and market competition)
   - Commercial barriers

C. Applicability to the UK
   - Markets and sectors
   - Geography, raw materials and other physical factors
   - Alignment with UK suppliers and R&D efforts/programmes

# Results of assessment

| CCU category | Technology/application | CRITERIA | | | Selection? |
|---|---|---|---|---|---|
| | | A. Technology development and performance | B. Economic and commercial potential | C. Applicability to the UK | |
| $CO_2$ to fuels | Renewable methanol and methane production | TRL 4-8 | | | YES |
| | Formic acid production | TRL 5 | | | NO |
| | Algae cultivation | TRL 3-5 | | | NO |
| | Helioculture | TRL 3 | | | NO |
| | Counter Rotating Ring Receiver Reactor Recuperator | TRL 3 | | | NO |
| | Photocatalytic reduction of $CO_2$ (metallic) | TRL 3 | | | NO |
| | Photocatalytic reduction of $CO_2$ (non-metallic) | TRL 3 | | | NO |
| | Nanomaterial catalysts | TRL 2-3 | | | NO |
| Enhanced commodity production | Enhanced Geothermal System with $CO_2$ | TRL 4 | | | NO |
| | Supercritical $CO_2$ power cycles | TRL 3 | | | NO |
| | Urea yield boosting | TRL 9 | | | NO |
| | Methanol yield boosting (conventional) | TRL 9 | | | NO |
| $CO_2$ mineralisation | Mineral carbonation | TRL 3-7 | | | YES |
| | Sodium bicarbonate | TRL 6 | | | NO |
| | $CO_2$ concrete curing | TRL 5 | | | NO |
| | Bauxite residue carbonation | TRL 8 | | | NO |
| $CO_2$ as chemicals feedstock | Polymer processing (polycarbonates) | TRL 3-5 | | | YES |
| | Polymer processing (polyurethanes) | TRL 3-5 | | | YES |
| Other existing commercial applications | Food and beverage applications | TRL 9 | | | YES |
| | Horticulture | TRL 9 | | | YES |
| | Other Industrial and technical uses | TRL 9 | | | YES |

# Renewable methanol ✔

- Brief description: Electrolysis of water to produce hydrogen, which is then combined with $CO_2$, compressed and reacted over a catalyst to produce methanol and water. Methanol can be blended with gasoline into various grades of transport fuel. Energy provided by renewable energy source offers potential for low-carbon fossil fuel substitution combined with renewable energy storage. Various process routes and hydrocarbon products via syngas can be achieved e.g. MBE.
- Technology status: Currently operating on commercial scale in Iceland, albeit under specific circumstances (surplus renewable energy + high fuel import prices).
- Technology providers/R&D efforts: Carbon Recycling International (Iceland); Haldor Topsoe (Denmark); various R&D programmes worldwide/UK into use of catalysts
- Economic and market factors: Economics are highly dependent upon relative costs of renewable energy source and (conventional) fossil-based transport fuel. Support for and regulation of alternative transport fuels (EU & UK level) is a key market factor
- Key barriers and challenges: Developing markets for methanol; ongoing reduction of capital costs through scale-up and increasing process efficiency
- UK perspective: Likely to be of most potential applicability to UK, in view of e.g. very large challenges to hydrogen energy (e.g. via $CO_2$ to formic acid) and algae (limited role in UK). Potential link to UK offshore wind strategy etc.

# Mineral carbonation ✔

- Brief description: $CO_2$ is reacted with minerals - mostly calcium or magnesium silicates - to form (Ca or Mg) carbonates (e.g. limestone) for use in building materials with storage of industrial $CO_2$. Unlike with other uses of $CO_2$, the process can work directly from flue gas (i.e. no capture step required).
- Technology status: Various process routes, all currently at pre-commercial stage
- Technology providers/R&D efforts: Calera (USA), Skyonic Corporation (USA), Bechtel (USA), Capitol Aggegrates Ltd (USA), Polarcus (Global), Novacem (now aquired by Calix) (UK) Cambridge Carbon Capture (UK), University of Sheffield (UK), Åbo Akademi University (Finland); Innovation concepts BV (Netherlands); Carbon-8 (UK);
- Economic and market factors: Existing market demand for low-carbon building products, subject to meeting regulatory/standard product reqiurements
- Key barriers and challenges: Achieving acceptable carbonation reaction rates remains key challenge to commercial scale-up
- UK perspective: Significant UK activity and industry collaboration; ETI study indicates UK has significant mineral deposits for commercial scale productions; use of industrial waste in manufacture has been demonstrated in UK e.g. Carbon-8

# Polymer processing ✔

- Brief description: Use of captured $CO_2$ in combination with traditional feedstocks to synthesise polymers such as polypropylene carbonate (PPC) and polyethylene carbonate (PEC) for use in various products and applications.$CO_2$ can also be used as a feedstock in the polymerisation of urethanes to produce polyurethanes.
- Technology status: Remains at pre-commercial stage with only small-scale demonstration to date (using a batch reactor). Significant industry involvement with first commercial applications expected within the next five years.
- Technology providers/R&D efforts: Industry: Bayer DREAM project (Germany), BASF (Germany), RWE (Germany); R&D efforts incl: Coates Group, part of Cornell University (USA), Novomer Ltd (USA), Kodak Speciality Chemicals (USA), Praxair (USA), Albermarle Corporation (USA) and Eastman Kodak (USA)
- Economic and market factors: Costs remain prohibitive compared to conventional polymers; significant market size globally with strong growth outlook
- Key barriers and challenges: Costs and existing polymer products on market
- UK perspective: Major funding is currently in Germany and the US; however, various ongoing R&D activities exist within UK, UK has large chemicals knowledge capacity, and existing efforts (e.g. DREAM) look to expand to other regions/sites

# Existing commercial uses of $CO_2$ ✔

- Brief description: in addition to EOR and urea manufacture, $CO_2$ is currently used across a wide range of smaller-scale sectors and applications including food and beverages, horticulture, pharmaceuticals, pulp and paper processing, water treatment, steel manufacture, electronics, pneumatics and welding. $CO_2$ is also used as a refrigerant gas and for fire suppression.
- Technology status: Established commercial usage across a range of sectors, either to industrial grade (>99% conc.) or food grade (>99.9% conc.).
- Technology providers/R&D efforts: Product $CO_2$ provided by large range of companies including e.g. Linde/BOC, Air Liquide, Air Products, Praxair, Messer.
- Economic and market factors: Costs of production highly dependent upon process/source, production volumes and purity requirements. Established market limited in size, and with existing market suppliers.
- Key barriers and challenges: Costs of capture and delivery; existing market players.
- UK perspective: There is demand for $CO_2$ across various sectors and sites, albeit with limited overall volumes. Some latent demand for close-proximity $CO_2$ supply may exist in some local industrial settings.

# CCU Stakeholder consultation

### Stakeholder consultation on CCU aspects of study:

- Views, information and feedback sought through (b) interim results workshop/meeting; (b) CCU questionnaire; and (c) follow-up communications

### Questionnaire sent to around 20 stakeholders (CCU tech providers, industry, academia)

- Feedback sought on the CCU technology ‘long list’ and assessment criteria
- Views on the potential for CCU in the UK; what are the drivers? which sectors? etc
- What are the key challenges and opportunities for CCU deployment?

### Results of consultation

- Broad support for the technology ‘long list’, the assessment approach criteria, and the resulting ‘short list’; renewable methane production added, following feedback from 2 stakeholders
- General view that CCU had some market potential within the next decade, with greatest potential over the longer term
- The role of renewable energy, and its potential alignment with some CCU technologies, was highlighted in several responses

# CCU Stakeholder consultation

Stakeholders asked to rank obstacles to CCU deployment in UK (3 = critical; 1 = weak):

| Obstacle | Response A | Response B | Response C | Response D | GROUP AV. |
|---|---|---|---|---|---|
| Cost factors | 3 | 3 | 2 | 3 | **2.75** |
| Insufficient support/incentives | 3 | 2 | 3 | 2 | **2.5** |
| Regulatory uncertainty | 1 | 2 | 1 | 3 | **1.75** |
| Low understanding of $CO_2$ utilisation technology | 1 | 1 | 2 | 2 | **1.5** |
| Unproven technology | 2 | 3 | 1 | 2 | **2** |
| Inability to demonstrate GHG benefits | 1 | 3 | 1 | 1 | **1.5** |
| Insufficient technical skills and know-how | 2 | 2 | 2 | 2 | **2** |
| Low suitability to UK industry/sectors | 2 | 2 | 1 | 1 | **1.5** |
| Undeveloped markets for $CO_2$-using products | 3 | 2 | 1 | 3 | **2.25** |
| High energy (or input) requirements | 2 | 3 | 2 | 2 | **2.25** |
| Integration within existing industrial processes | 3 | 2 | 1 | 3 | **2.25** |

- Broad consensus that cost factors and lack of incentives were key obstacles
- In general, UK applicability and understanding of technology not seen as obstacles
- Less consensus around market and regulatory factors; mixed views

# CCU Deployment scenarios

- Illustrative scenarios of CCU deployment in the UK developed for the selected applications/technologies (the short list)
- Objective: to describe, at a high level, a viable range of CCU deployment in 2025 in terms of industrial $CO_2$ utilised (million tonnes $CO_2$ per year) and potential revenues from CCU products (million £ per year)
- CCU technology modelled against the data for UK industrial $CO_2$ sources; illustrative CCU products chosen (a wider range of competing processes and products in reality)
- Three scenarios (*see next slide*) then present three progressively ambitious outlooks, or pathways, for UK uptake of CCU technology through 2025…
- **Very high** scenario can be considered at the very upper end of what would be feasible by 2025, given the current low-zero level of deployment within the UK (other than small-scale R&D lab and pilot efforts)
- Even the **moderate** scenario would entail significant technology progress, policy support and/or favourable market development for CCU products over the next decade; should therefore not to be interpreted as a 'business as usual' type scenario
- **High scenario** is considered to be illustrative of what could be achievable given significant support environment for CCU, both in terms of *market push factors* (e.g. regulatory and financial support) and *market pull factors* (e.g. demand for CCU products, product price increases, carbon pricing incentives)
- Note that the low scenario, corresponding to negligible CCU, is not shown.

# CCU Deployment scenarios

| CCU uptake in 2025 | CCU application | | | | |
|---|---|---|---|---|---|
| | **Renewable methanol** | **Renewable methane** | **Mineral carbonation** | **Polycarbonates** | **Industrial product $CO_2$** |
| **Very high** | 10% penetration of the UK road transport petroleum market in 2025. Equivalent to around 2.75 million tonnes annual methanol production (roughly one third of current methanol fuel blending globally, and around 4-7 commercial scale plants) using around 3.75 million $tCO_2$ p.a. | 5% penetration of the UK natural gas power generation market (on an energy basis). Equivalent to around 11 TWh (approx. 1 billion $m^3$) annual methane production and around 1.8 million $tCO_2$ utilisation. | Up to 50% of the UK's cement sector emissions used for mineral carbonation products. 10% of magnesite production [3] is used in early-stage high value industrial applications; 90% is used in lower value bulk markets such as lime, filler etc. Up to 10% of bi-product APS production potential realised.[4] | 1-2 commercial-scale plants by 2025 with capacity of approx. 300,000 tonnes p.a. PEC or PPT, utilising industrial $CO_2$ of approx. 150,000 $tCO_2$ p.a. Represents just 4% of the current PE market in Europe - although considerable obstacles face development of PEC production, and investor confidence, within the UK. | Assumes 20% market growth through 2015-2025, based on estimated current demand of 200,000-300,000 $tCO_2$ p.a. (i.e. additional demand of 50,000 $tCO_2$ p.a. across a range of sectors and applications e.g. beverages, horticulture, electronics, waste water, speciality chemicals. |
| **High** | 5% penetration of the UK road transport petroleum market. Equivalent to around 1.37 million tonnes annual methanol production and around 1.9 million $tCO_2$ utilisation p.a. (2-4 commercial scale plants). | 1 commercial-scale plant operational by 2025 with production capacity of approx. 15 million $m^3$ utilising industrial $CO_2$ from on-site or other nearby $CO_2$ sources of approx. 30,000 tCO2 p.a. | Up to 25% of the UK's cement sector emissions utilised. 5% of magnesite production is used in early-stage high value industrial applications and only 5% of APS production potential is able to find a market (60-70,000 t). | 1 commercial-scale plant by 2025 with production capacity of approx. 100,000 tonnes p.a. PEC or PPT, utilising industrial $CO_2$ of approx. 50,000 $tCO_2$ p.a. Represents less than 2% of the current European PE market. | 10% market growth through 2015-2025 i.e. additional demand of 25,000 $tCO_2$ p.a. across a range of sectors and applications. |
| **Moderate** | 1 commercial-scale plant operational in the UK by 2025 with capacity of approx. 50 million litres utilising industrial $CO_2$ from on-site or other nearby $CO_2$ sources of approx. 55,000 $tCO_2$ p.a.[1] | Pilot scale pre-commercial production only (100,000 $m^3$ methane p.a., equal to the world's current largest pilot project in Germany) supplied by c.200 $tCO_2$ p.a. [2] | Up to 10% of the UK's cement sector emissions utilised - equivalent to one typically sized cement plant of 0.6 $MtCO_2$ per year. All magnesite production is used in bulk applications and APS production is unable to find a market outlet. | Pilot scale pre-commercial production only (e.g. 10,000 tonne product p.a.) supplied by 5,000 $tCO_2$ p.a. Could operate as an R&D supported slip-stream $CO_2$ source within a larger CCS project including geological storage. | 5% market growth through 2015-2025 i.e. additional demand of 10-15,000 $tCO_2$ p.a. across a range of sectors and applications. |

Low scenario not shown – would involve negligible CCU. Note there is no assumed correlation with the $CO_2$ capture technology development scenarios in the period to 2025.

# Scenario modelling assumptions: Renewable methanol

- Process description: Electrolysis of water to produce $H_2$, with subsequent catalytic conversion (~5MPa, ~225$^{o}$C) of $H_2$ and $CO_2$ to methanol ($CH_3OH$) and water. For low-carbon fuel production (note that $CO_2$ is released upon fuel use), the process energy source would need to be from a renewable source.
- Product markets: Methanol blended with petrol for use as transport fuel. Current global production of methanol is around 40Mtpa (IMPCA, 2013); significant growth forecast with potential for M15 blends using as per China. However, methanol blending currently limited to 3% in EU: higher rates only allowed for bio-methanol under the RE Directive
- Product price range: European Methanol prices (FOB Rotterdam) were EUR 390/tonne in Q2 and Q3 2013, having varied from around EUR 150-500/tonne over the past 5 years (INEOS, 2014). Prices are forecasts to grow through 2014 with strong demand globally, particularly in Asia (ICIS, 2014). Range of <u>EUR 300-500/tonne</u> chosen.
- $CO_2$ utilisation rate: Methanol synthesis from $CO_2$ and $H_2$ (from electrolysis of water) converts 1 $tCO_2$ captured into (32.04g/mol)(44.01g/mol) t of methanol ($CH_3OH$) i.e. <u>0.728 t methanol/$tCO_2$</u>. Or, 1 tonne methanol requires 1.374 $tCO_2$ input.

# Scenario modelling assumptions: Renewable methane

- Process description: The Sabatier reaction exothermically combines hydrogen and carbon dioxide to produce methane and water [$CO_2 + 4H_2 = CH_4 + 2H_2O$]. The reaction is usually carried out in the presence of a nickel catalyst. For low-carbon fuel production (note that $CO_2$ is released upon fuel use), the process energy source would need to be renewable.
- Product markets: Methane used in energy supply e.g. power generation, domestic and industrial applications. Strong demand for gaseous fuels forecast within UK, the EU and globally.
- Product price range: UK wholesale natural gas prices between 2007-2013 have trended between around 40 and 80 pence/therm. Future pricing will be determined by a range of unknown factors impacting regional and UK wholesale gas markets. DECC forecast UK gas prices in 2025 to be between 42.2 (low) and 105.4 (high) p/therm, with a central estimate of 73.8 p/therm (DECC, 2013). This range therefore chosen.
- $CO_2$ utilisation rate: Methane synthesis from $CO_2$ and $H_2$ (from electrolysis of water) converts 1 t$CO_2$ captured into (16.04 g/mol)(44.01g/mol) t of methane i.e. <u>0.364 t methane/t$CO_2$</u>

# Scenario modelling assumptions: Mineral carbonation

- Process description: Conversion of magnesium silicate (olivine & serpentine) into brucite ($Mg(OH)_2$) and amorphous precipitated silica (APS). The brucite can then be used as an agent to store flue gas $CO_2$ in the form of solid magnesite ($MgCO_3$).[(1)]
- Product markets: (1) Magnesite: has industrial uses in dry powder form e.g. as anti-caking agent and fire retardant. Market for these high value applications is growing but limited to a few million tonnes p.a. globally. Larger bulk markets are as lime and as construction aggregates/fillers (competing here with limestone and dolomite, but with a market volume of several billion tonnes p.a. globally). (2) APS: High value product used as a rubber filler for tyres. Market size approx. 2 million tonnes p.a.
- Product price range: Value for high grade synthetic magnesite USD 500-1000; Lower value larger market applications in the range USD 10-300 e.g. cementitious material to blend with Portland cement, or for use as a custom binder material, or as aggregate, or a soil amendment (CSLF, 2012); APS sells at USD 500-1000 per tonne
- $CO_2$ utilisation rate: Based on serpentine, 1t$CO_2$ sequestered produces 1.916 t magnesite (based on 1.4 t magnesite per tonne brucite) and 0.91 APS; Based on olivine, 1t$CO_2$ sequestered produces 1.916 t magnesite (based on 1.4 t magnesite per tonne brucite) and 0.81 APS.

(1) data and assumptions based on process and market information provided by Cambridge Capture Company

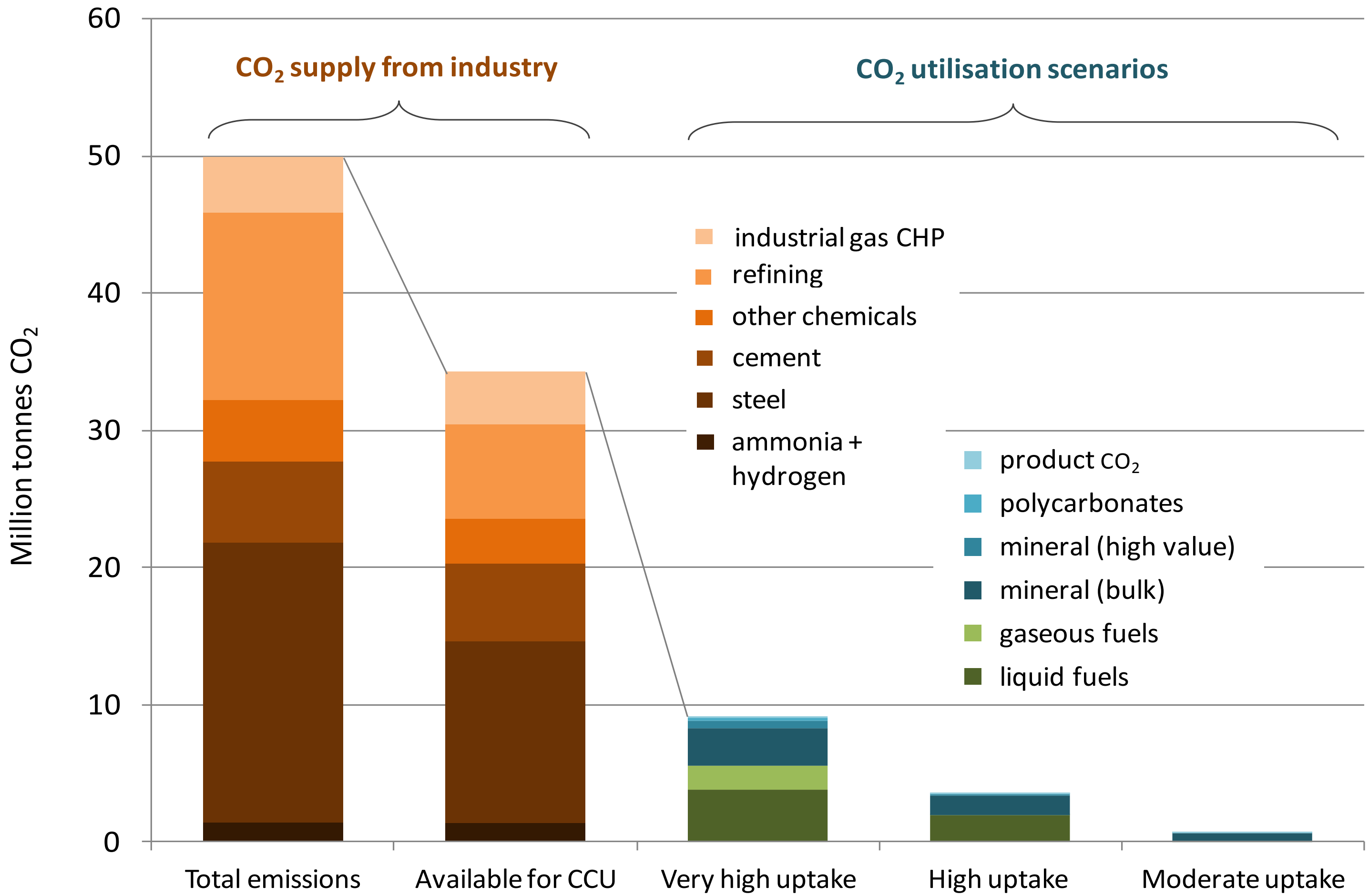
# Scenario modelling assumptions: Polycarbonates

- Process description: $CO_2$ used as feedstock for production of polycarbonates such as polypropylene carbonate and polyethylene carbonate (PEC), using a zinc-based catalyst in a reaction with epoxide molecules e.g. Novomer, US. (N.b. a variety of other process routes and end products have been proposed; Bayer has piloted production of polyether polycarbonate-polyols (PPP) used in production of high-grade plastic polyurethane).
- Product markets: Polymer coatings, plastic bags, laminates / coatings, surfactants for EOR, automotive and medical components. PEC market size in the U.S. 5 million tonnes/year (CSLF, 2012). The PEC market is of a similar size in the EU (PE = 17% of 47 Mt in 2012 according to PlasticEurope, 2012) . PE production is forecast to rise to 125 Mt globally in 2025 (ICIS, 2013).
- Product price range: PEC price is around USD 1,000/tonne (CSLF, 2012); PE prices have since risen to as high as EUR 1,500/t in Europe (ICIS, 2014).
- $CO_2$ utilisation rate: 1 $tCO_2$ produces 2.32 t polypropylene carbonate based on Novomer polymers case study in GCCSI (2011).

# Scenario modelling assumptions: Industrial product $CO_2$

- Process description: $CO_2$ sourced from various natural and industrial processes (e.g. ammonia plants, refineries, breweries, natural wells, and combustion sources) and subject to clean up to industrial grade (>99% $CO_2$ by volume or food grade (>99.9% $CO_2$ by volume).
- Product markets: Wide range of existing commercial applications across many industrial sectors including: food and beverages, horticulture, pharmaceuticals, pulp and paper, water treatment, electronics, fire suppression etc. Product $CO_2$ sold in various quantities and forms from small bottles/canisters on the retail market to larger bulk supply through bi-lateral contracts; provided by large range of companies including e.g. Linde, Air Liquide, Praxair, Messer, Yara. The market is limited and saturated by existing providers. Over 1Mt$CO_2$ is theoretically available in the UK from high purity sources, but market understood to be mature with little/no latent demand.
- Product price range: Typical market price of treated $CO_2$ can range from EUR 60-300/tonne according to region, application, volume sold etc (Find, 2013).
- $CO_2$ utilisation rate: N/A

# Annual UK industrial $CO_2$ supply and utilisation potential in 2025



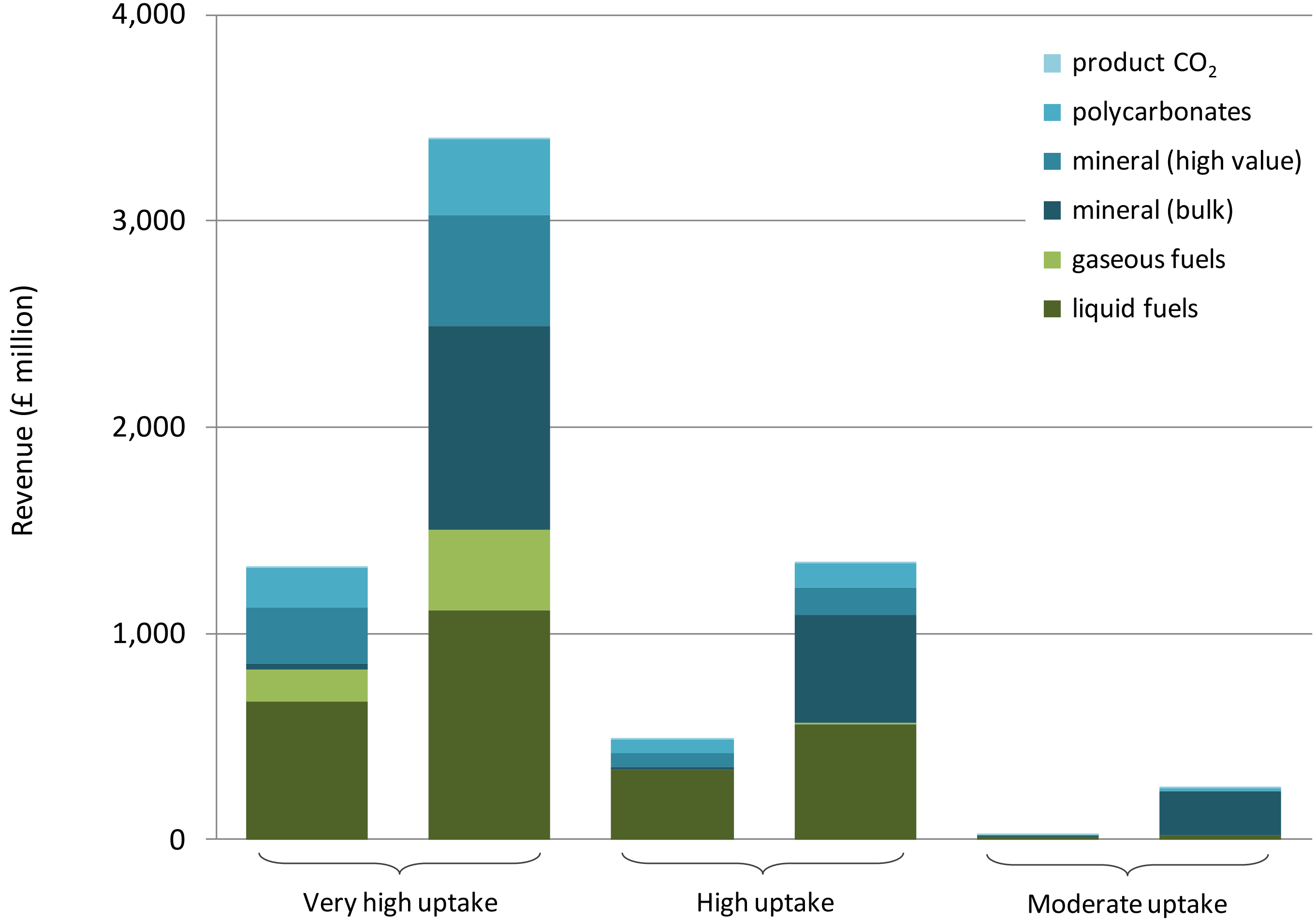

- **Very high** scenario: 8-9 million $tCO_2$ utilised per year =15-20% of UK industrial emissions (or all chem. Industry $CO_2$)
- **High** scenario: falls to 3-4 million $tCO_2$ (approx. 7% of total emissions)
- **Moderate** scenario: around 0.5-0.7 million $tCO_2$ (approx. 1% of total emissions)
- **Low** scenario: *ca*. 0 $MtCO_2$/yr (not shown)

# Annual UK industrial $CO_2$ supply and utilisation potential 2025

| *$MtCO_2$* | Total emissions | Available for CCU |
|---|---|---|
| Cement | 6.0 | 5.7 |
| I&S | 20.4 | 13.3 |
| Refining | 13.7 | 6.9 |
| Chemicals (pure CO2) | 1.4 | 1.3 |
| Chemicals (CHP) | 4.0 | 3.8 |
| Chemicals (other) | 4.4 | 3.3 |
| **TOTAL** | **49.9** | **34.3** |

| *$MtCO_2$* | Very high uptake | High uptake | Moderate uptake |
|---|---|---|---|
| Liquid fuels | 3.8 | 1.9 | 0.0 |
| Gaseous fuels | 1.8 | 0.0 | 0.0 |
| Mineral (bulk) | 2.7 | 1.4 | 0.6 |
| Mineral (high value) | 0.6 | 0.1 | 0.0 |
| Polycarbonates | 0.2 | 0.1 | 0.0 |
| Product CO2 | 0.0 | 0.0 | 0.0 |
| **TOTAL** | **9.0** | **3.5** | **0.6** |

# **Annual revenue potential (market size) 2025** - low and high ranges




- **Very high** scenario: £1.3-3.4 billion
- **High** scenario: £0.5-1.3 billion
- **Moderate** scenario: £25-250 million
- **Low** scenario: £0 million (not shown)

# **Example of estimated revenue calculation** – renewable methanol under ‘high’ CCU deployment scenario

- Under high scenario, assumes 10% market penetration by 2025 of current UK demand for motor spirit on energy basis (i.e. 10% * 13.23 million tonnes motor spirit in 2012 according to *Digest of UK Energy Statistics 2013*) * (32.4 MJ/litre motor spirit / 15.6 MJ/litre methanol)
- Product potential in 2025 therefore = <u>2.75 million tonnes methanol</u>
- Product price assumed = <u>300-500 EUR/tonne</u> *(see earlier slides)*
- Exchange rate (GBP:EUR) = <u>1.233</u> *(DECC September 2013 version of the appraisal guidance)*
- Therefore, potential revenue range estimated = (2.75*300/1.233)-(2.75*500/1.233) = <u>£669-1,114 million per year</u>
- In a low CCU scenario, there would be no revenues from $CO_2$ utilisation.

# **Annual revenue potential (market size) 2025** - low and high ranges

| | Very high uptake | | High uptake | | Moderate uptake | |
|---|---|---|---|---|---|---|
| ***£ million*** | low | high | low | high | low | high |
| Liquid fuels | 668.6 | 1114.3 | 334.3 | 557.2 | 9.7 | 16.2 |
| Gaseous fuels | 153.7 | 384.0 | 2.3 | 5.8 | 0.0 | 0.0 |
| Mineral (bulk) | 33.1 | 992.5 | 17.5 | 523.8 | 7.4 | 220.6 |
| Mineral (high value) | 269.1 | 538.3 | 67.3 | 134.6 | 0.0 | 0.0 |
| Polycarbonates | 189.0 | 365.0 | 63.0 | 121.7 | 6.3 | 12.2 |
| Product CO2 | 2.4 | 12.2 | 1.2 | 6.1 | 0.6 | 3.0 |
| **TOTAL** | **1316.0** | **3406.3** | **485.6** | **1349.1** | **24.0** | **252.0** |

- Potential revenues/value from $CO_2$ avoidance not included in analysis
- Net GHG benefits have not been assessed: further LCA-type assessment is needed based upon specific technologies/applications with e.g. defined scope and boundaries
- Costs have not been assessed: robust analysis would similarly require technology and setting-specific assessment
- Low scenario not shown, but would involve no additional uptake of $CO_2$ utilisation.

# Key references

## Renewable methanol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- Any conclusions represent the views the authors alone and not those of DECC, BIS, the project industrial steering board, or the consultees.

# Caveat

- While the authors consider that the data and opinions contained in this report are sound, all parties must rely upon their own skill and judgement when using it.
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- All databases and models discussed in this document involve many assumptions and necessary simplifications of a complex reality.
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