



Programme Area: Carbon Capture and Storage

Project: Mineralisation

Title: Carbon Capture and Sequestration by Mineralisation (CCSM) Stage 2A Report

Abstract:

This report describes the state of technology development and compares CCS using Mineralisation (CCSM) with Geological CCS (GCCS). It concludes by describing a programme of experimental work and economic analysis that will identify opportunities to improve the competitiveness and operability of CCSM processes. This report should be read in conjunction with the Stage 2a Techno-Economic Analysis (TEA) Report Number 11. The economic feasibility of CCSM relative to other CO₂ capture processes is studied through capital cost and operating cost. This report builds on the knowledge derived from the initial techno-economic assessment and by scrutinising the cost structure and key attributes of CCSM technology building blocks available in the open literature. As well as addressing the challenges highlighted in this report and TEA-1 it also looks at data gaps from Work Package 1. Results from the experimental plan will create the dataset required for the broader and more detailed Techno-Economic Assessment of selected CCSM technologies. The ETI is grateful for the contributions made by all participants of the Mineralisation project; Perkins Engines Company Limited, Shell Global Solutions International B.V., Natural Environment Research Council as represented by the British Geological Survey, and the University of Nottingham.

Context:

CCS by mineralisation has been identified as a promising additional method of sequestering CO₂ emissions. Minerals and CO₂ can react together to permanently store CO₂ as a solid carbonate product, which can then be safely stored, used as an aggregate or turned into useful end products such as bricks or filler for concrete. This £1m project, launched in May 2010 carried out a detailed study of the availability and distribution of suitable minerals across the UK along with studying the technologies that could be used to economically capture and store CO₂ emissions. The project consortium involved Caterpillar, BGS and the University of Nottingham. The objective was to investigate the potential for CCS Mineralisation to mitigate at least 2% of current UK CO₂ emissions and 2% of worldwide emissions over a 100- year period. The project has found that there is an abundance of suitable minerals available in the UK and worldwide to meet these mitigation targets. However, challenges remain to make the capture process economically attractive and to reduce its energy use. Significant niche opportunities exist where waste materials are used as feedstock and/or the process produces value-added products, but markets would not be at the level required to meet the mitigation targets.

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Carbon Capture and Sequestration by Mineralisation (CCSM) Stage 2A Report

CCSM Programme

Stage Gate Report

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ENERGY TECHNOLOGIES INSTITUTE

CCSM PROGRAMME

STAGE 2A REPORT

Carbon Capture and Sequestration by Mineralisation (CCSM) Stage 2A Report

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1 Introduction

This report outlines the position of Carbon Capture and Storage using Mineralisation (CCSM). It describes the state of technology development and compares CCSM with Geological Carbon Capture and Storage (GCCS). It concludes by describing a programme of experimental work and economic analysis that will identify opportunities to improve the competitiveness and operability of CCSM processes.

This report should be read in conjunction with the Stage 2a Techno-Economic Analysis (TEA) Report Number 1¹ also prepared for the Energy Technologies Institute (ETI). This report adds to the TEA-1 and discusses the proposed forward work programme in detail. The economic feasibility of CCSM relative to other CO₂ capture processes is studied through capital cost (CAPEX) and operating cost (OPEX). A simple excel based CAPEX analysis model has been constructed using published data and chemical engineering standard references. The tool allows the team to estimate the high-level impact of changes in each CCSM process component. It can also be used to test cost reduction opportunities.

This report builds on the knowledge derived from the initial techno-economic assessment and by scrutinising the cost structure and key attributes of CCSM technology building blocks available in the open literature. As well as addressing the challenges highlighted in this report and TEA-1 it also looks at data gaps from Work Package 1. Results from the experimental plan will create the dataset required for the broader and more detailed Techno-Economic Assessment of selected CCSM technologies.

¹ TEA-1, CPI for the ETI CCSM Constortium, July 2011

2 Capital Costs of CCSM technologies

2.1 Introduction

The TEA-1 report focused on the Shell Global Solutions (SGS) process, and this data is summarised here. A wide range of CCSM costs have also been published by various authors for various processes. Only two detailed economic analyses have been published in the open literature. They are from the Energy Centre of the Netherlands (ECN, [1]), which published an economic analysis describing a “direct (single-step), aqueous process with no additives”. In this report, the CCSM CAPEX for a 100MW power plant is calculated to be £40M, and £90/tonne CO₂ was estimated. The other data comes from Albany Research Centre (ARC) / National Energy Technology Laboratory (NETL) economic report [2] describes a “multi-step, aqueous, additive-enhanced” CCSM technology; the associated CAPEX for 1GW power plant was £998M, and the estimated sequestration cost was £123/tonne CO₂. In addition, Shell cost data has been made available to the consortium. Capital cost to build a Shell CCSM Plant for a 170MW power plant is £720M with a capture cost (without capital costs) of £155/t of CO₂ avoided. These data indicate that the published data vary widely, and it is not clear what has been included in many of the analyses, making direct comparison of the results difficult. It is not clear whether the difference in reported costs (in £/tonne CO₂) is due to fundamental differences in the process efficiency or purely due to the economic models and assumptions used.

The CAPEX software tool (CCSM-CAPEX) has been created to provide a high level and comparative analysis of the CAPEX of various CCSM technologies. The tool is based on an Excel worksheet. The Generation 1 development includes data on capital equipment costs for two CCSM technologies obtained from the open literature: “Aqueous, single-step, no additives” (ECN, CAT) and “Aqueous, multi-step, additive-enhanced” (ARC/NETL). The model is still under development and will be used in the next stages of the project to look at comparative economics of processes.

2.2 Capital Expenditure Analysis (D2.4vi)

The analysis of capital costs data was performed to:

- Compare the Purchased Capital Equipment (PCE) cost structure for each technology.
- Identify the unique capital equipment components and common components of the various technologies.
- Define the most expensive (capital) units in each technology and the parameters that control their cost.
- Define the impact of differences in the cost models.

The following CCSM process input characteristics are used for the PCE estimation:

- CO₂ emission amount.
- Assumed or experimentally demonstrated CO₂ reduction efficiency.
- Type of and amount of minerals required per unit mass of CO₂.
- Additive(s) amounts per unit mass of mineral (or unit mass of CO₂).
- Liquid/solid ratio (assuming water).
- Chemical activation process parameters (temperature, pressure, time) - if applicable.
- Carbonation process parameters (temperature, pressure, time).
- Cost formulas obtained from the open literature for industrial equipment [4].

The analysis shows that the following equipment unit is common for all technologies:

- Mineral grinding equipment (crashes, ball mills)
- Aqueous slurry pumps with electro-motors
- Heat exchanges/coolers/heaters
- Tanks/vessels (for slurry hold and/or phase separation)
- Fans, blowers

A number of the common units account for a relatively low share of total PCE cost. For example, the combined cost contribution for crushers, pumps, motors, tanks, fans, blowers was approximately 13% PCE of ECN CCSM (Figure 1). It is not yet clear whether technology developments in recent years offer opportunities for improvement and this will be reviewed at the next stage of the project. In addition it is worth noting that if the operating conditions of the main reactors are improved then there is a good chance that the equipment listed here will also get smaller and hence have a lower capital requirement.

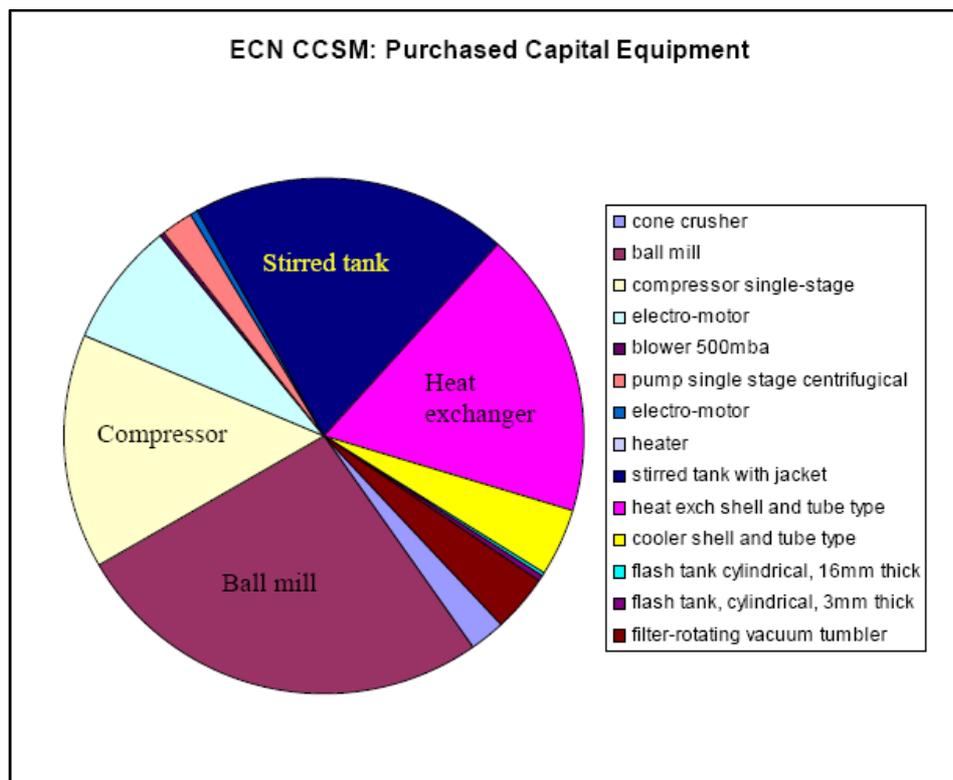


Figure 1. PCE Cost Structure for ECN CCSM.

However, the physical challenge of handling such large volumes of effluent gas has the knock on effect of creating a large-scale plant, which by definition has large capital costs associated with it. The following bullet point list demonstrates the challenge:

- As a rule of thumb gas fired power stations have a capital cost of about £500k per MW²;
- Therefore a 300 MW gas fired power station would be a £160 million capital investment;
- This would produce 1,000,000 tonnes of CO₂ per year;
- According to Shell Global Solutions¹ (SGS) a carbon dioxide mineralisation facility which would capture 50% of the power station's carbon dioxide emissions would cost of the order of £720 million capital.

² Based on US EIA Annual Energy Outlook 2007 converted for UK.

The capture plant proposed by SGS is limited to 50% CO₂ avoidance by its design and operation. The CCS plant captures 80% of the CO₂ emissions from the power station, but the high temperature thermal activation step uses large amounts of imported energy to reach its operating temperature and the end result is that the carbon dioxide avoided after this imported fossil energy is taken into account is only 50% of the power station emission.

This indicates that the cost of the mineralisation plant, based on the initial Shell plant design created by SGS would cost four times as much as the power station to build and is likely to be up to four times larger than the power station itself. It is likely that on a comparable basis this will be similar ratio for other technologies.

The conclusion from the high level analysis is that the capital cost and scale of the plant need to be significantly reduced before becomes an attractive investment.

As stated above the largest amount of capital data has been provided by SGS for their thermally activated process. The process has been researched and developed over more than 8 years. As well as desk analysis, laboratory research and initial process testing (on a dedicated rig at ECN in the Netherlands) SGS has created an Aspen model of the process. However, work has recently been stopped and Shell has withdrawn from the ETI project. Their work has been made available to the project consortium as a full Shell Confidential report, and the Aspen model has also been made available. This is located on the Aspen platform at CPI and can be used for future process modelling. The outline of the Shell process is shown in Figure 2.

The main elements of the process are shown below and are described in detail in the TEA-1 report¹.

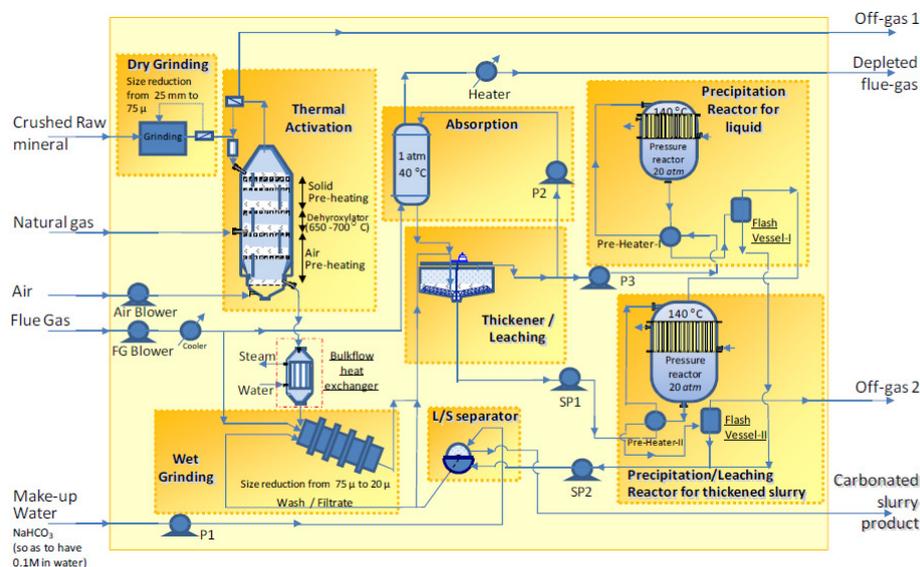


Figure 2. Outline of the Shell Mineralisation Process

The equipment costs for carbon capture and storage through mineralisation (CCSM) plant using the Shell process to capture 80% of the CO₂ from a power station producing 1 million tonnes of CO₂ would be £107 million leading to an installed cost of £720 million. This shows that SGS is

assuming an installation factor of 6.7. This is to the high end of the normal estimating processes, but as this is such a large complex plant with very substantial steel work, piping, control and civil requirements it is considered to be a reasonable estimate of the costs.

Figure 3 shows the split of costs between the major plant items.

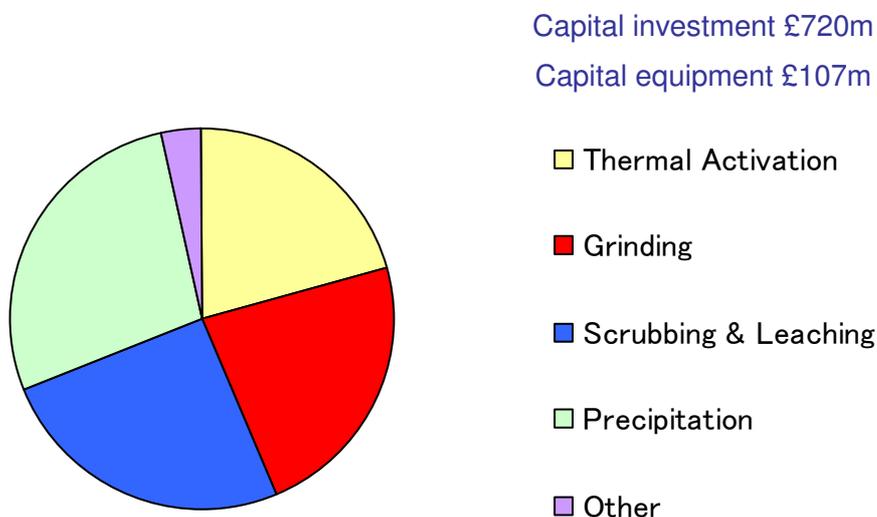


Figure 3. Main Plant Items - Shell Process

The figure shows that about 20% of the capital is spent on thermal activation and a further 20% on grinding processes. Scrubbing and leaching amounts for 25% of the cost and the precipitation units for about 30%.

The capital profile for the ECN plant quoted in the literature is shown in Figure 1. The ball mill expenditure for the ECN is 38% of PCE (ECN CCSM). The cost of a ball mill is controlled by the amount of mineral required per tonne CO₂ (or R_{CO2}). The additional thermal activation step (so-called dehydroxylation) in the case of serpentines (ARC CCSM and Shell) is very cost intensive and results in capital costs rising by about 21% and operating costs potentially rising further. Grinding equipment has been utilised in mineral engineering industry for a long time and has been largely optimised. The integration of mineral grinding and thermal activation steps within the overall mass and energy balance has not been addressed. Thermo-mechanical and mechano-chemical treatments are well-known, large-scale processes that have not been considered in application to CCSM. Therefore, the follow-on experiments should include this cost reduction opportunity.

The carbonation stage is also a complex process and differs across the cost models studied. For example, carbonation based on a pure CO₂ stream requires a combination of heat exchangers (up to ~18% ECN PCE, Figure 1), compressors with motors, and stirred tank reactors (~20% PCE for ECN, 62% PCE for ARC, Figure 4); a Spray Dry Absorber (SDA) is designed for flue gas treatment, but accounts up to ~74% PCE for CAT CCSM. The analysis showed that the CO₂ separation cost was not included into economics of ECN and ARC/NETL CCSM, while it could be significant contributor. CO₂ compressor cost is ~15% PCE for ECN technology; it is controlled by the required pressure for the follow up carbonation stage. Obviously, milder pressurisation would reduce costs.

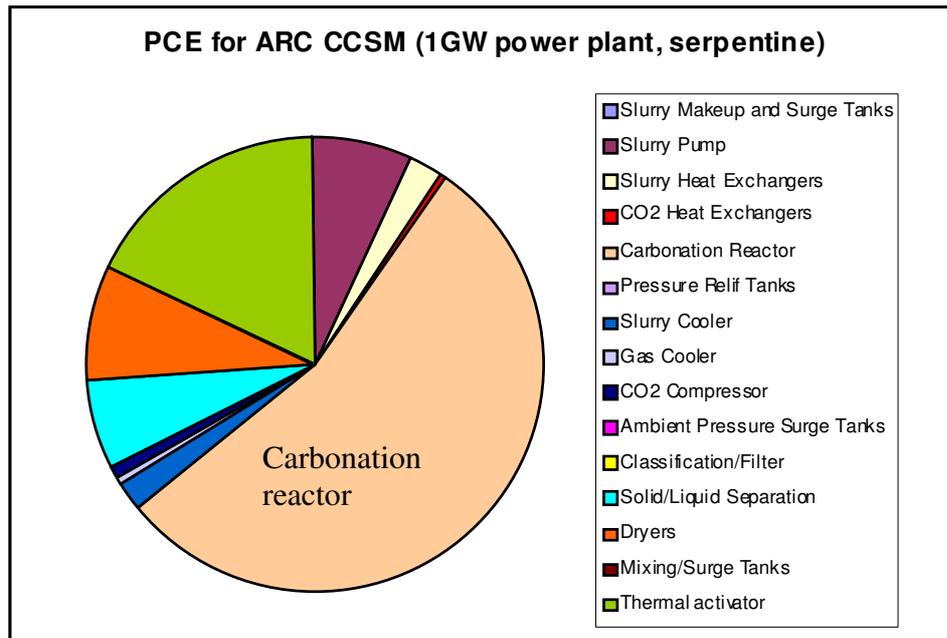


Figure 4. PCE structure for ARC CCSM: serpentine.

The cost of carbonation reactors (e.g. stirred tank reactor, SDA) is primarily controlled by balancing the requirements of residence and reaction time; therefore, faster carbonation kinetics lower cost. Carbonation temperature (T) and pressure (P) impact carbonation kinetics and the material choices for the reactor, but the materials choices are secondary parameters from the cost perspective. The carbonation reactors make up the bulk of the process investment in the ARC work, but it is not clear to what level the design was made. However, using a very simple treatment that is shown in Figure 5 it is found that the cost of the carbonation reactors for ARC/NETL would decrease from ~£620M to ~£310M (a factor of 50%) if the residence time in the reactor were halved (Figure 5a) or the pressure halved from 2200psi to 1100psi (Figure 5b). This indicates there is potential to reduce capital cost through better reaction understanding coupled with an improved design approach.

Spray dryer absorber costs are driven by the amount of liquid (water) that evaporates during the carbonation process; therefore, reducing the amount of water required will lower the SDA capital cost. Aqueous carbonation processes require large amounts of water and use stirred tank reactors in combination with flash tanks and filters (such as ARC/NETL processes). The SDA is more expensive than stirred tank reactors, but it includes an additional valuable function such as drying the final product and this was not considered in the other CCSM cost estimates. It is however, extremely difficult to manage the reactor in operation. The carbonation equipment is a major element of the PCE cost in all cases, and therefore improvements in reactor selection and design offer a significant cost reduction opportunity. It is recommended that the follow up experimental work should be focused on improving the kinetics of the carbonation step through mineral selection, pre-treatment selection, additive selections as well as large scale reactor design.

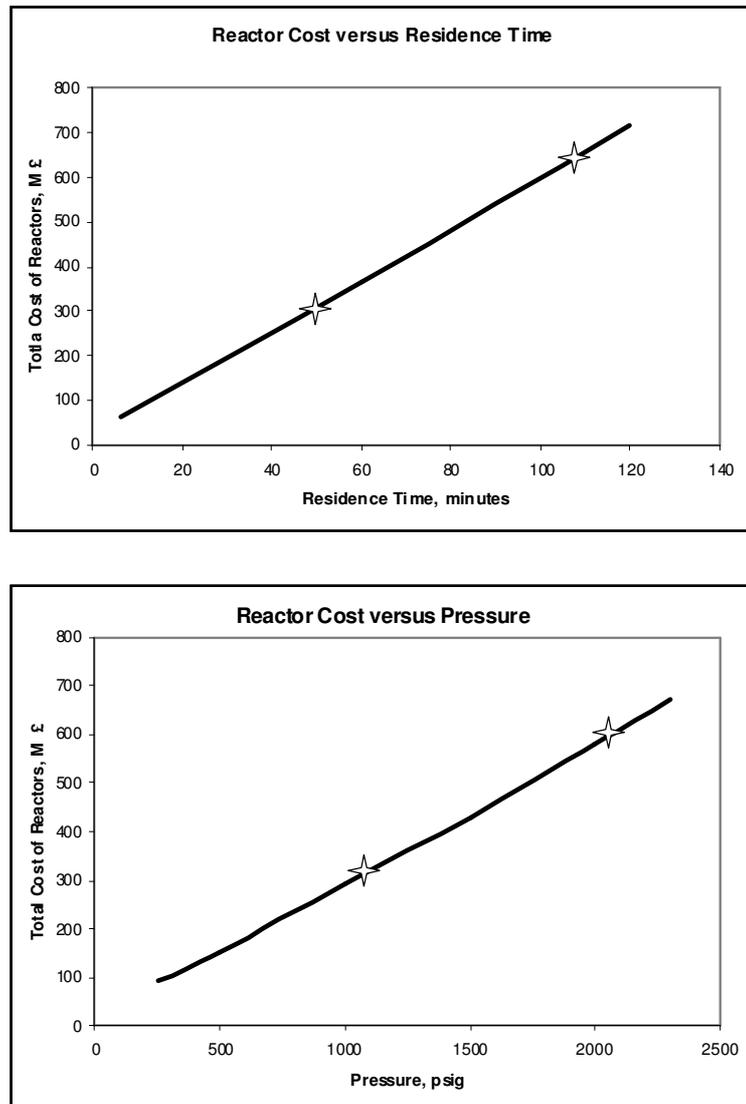


Figure 5. Sensitivity study for the carbonation reactor cost for ARC process: (a) impact of residence time; (b) impact of carbonation pressure. Sensitivity based on 95 reactor vessels of 378.5m³ able to treat 24,000tonne CO₂/day.

Analysis of CAPEX shows that different cost models were used by different authors. So direct comparison is difficult, but figures are typically between £50m and £100m for basic equipment. This would equate to an installed cost of between £350m and £700m for a 300MW plant with the Shell process at the higher end of the estimated range.

2.3 Conclusions

At this stage of development it is extremely difficult to compare the CAPEX of different CCSM technologies. As the project continues the consortium will update the Excel capital models to improve accuracy and comparability. Examples of parameters to be improved are:

- Same amount of CO₂ emissions should be selected for each technology
- Comparable basis for the CAPEX models used
- Choice of economic lifetime that is directly linked to the potential lifetime of the main plant equipment such as carbonation reactor(s), CO₂ compressors, thermal activation etc.

In addition the plant models will be improved to include:

1. Energy consumption for equipment operation. The energy consumption for equipment operation is a direct measure of CCSM energy penalty that should be considered while comparing various CCSM technologies.
2. Comparative cost analysis of various large-scale carbonation reactors. It was concluded above that selection and design of carbonation reactor(s) presents the significant area for cost reduction. Therefore, various chemical reactor designs (pipeline, recirculated/loop-type, stirred tank, SDA, etc.) will have to be evaluated for 1-2 selected CCSM technology(ies) from cost and energy penalty point of view.

In the Shell Aspen models that have been made available to the project team the capital and operating costs have been calculated. They arrived at a figure of £155 per tonne CO₂ avoided for the operating cost and a figure is £305 per tonne CO₂ avoided if the cost of capital is assumed to be 10%. Clearly at this amount the process will never be viable as it amounts to a cost of £1.45 per kWh of electricity (For reference current electricity price is around 13.5p/kWh).

The operating cost calculation includes raw material costs, utilities, people, overheads, and maintenance. The SGS report does not detail the distribution of costs between the operating activities. However, the report does state that the SGS technology CCSM plant requires 78MW of electricity to operate based on a power generation efficiency of 60%. This indicates that 25% of the power station output would be used to power the CCSM plant.

A parasitic electrical load of 25% indicates that considerable challenges in achieving the improvements required if an economically viable CCSM process is to be developed.

It is also worth noting that the capital cost of a flue gas desulphurisation plant that captures 90% of the sulphur dioxide released by a 330MW power station is £102 million. This is an order of magnitude smaller than the cost of the CCS mineralisation plant.

The conclusion from this analysis is that in order to create a more financially viable process significant improvements will be required in all of the major process elements of the CCSM process. The remainder of this report considers the rationale for the improvements and the work required to gather the data that will demonstrate if the improvement required is possible.

3 UK Scenario Analysis

CCSM is anticipated to have a different niche to Geologic Carbon Capture and Storage (GCCS). This section compares the current cost estimates to CCSM with GCCS. It is thought that CCSM will be most appropriate for (i) small to medium emitters on (ii) sites that are remote from geological storage sites. CCSM also needs (iii) access to large volumes of serpentinite rocks as sites are likely to be (iv) coastal locations.

3.1 Site selection

Two coastal sites have been chosen for this initial analysis. They are:

- Fellside Heat and Power plant on the coast of Cumbria, which will use rock from Ballantrae on the coast of Ayrshire. It produces 168MW heat and power, and 0.56Mtpa CO₂.
- Aberthaw Cement Plant in Glamorgan, South Wales, which will use rock from the Lizard in Cornwall. This site emits about 0.33Mtpa CO₂ per year.

The GCCS comparison is for storage in the North Sea, although it is recognised that there are potential storage sites in the Irish Sea.

The sites have been selected because they emit less than 1Mtpa CO₂, and they are of different types. The sites are located on the West coast of UK because these are reasonably distant from the current GCCS storage sites in the North Sea, and CCSM may be more applicable here. The sites are shown in Figure 6.

3.2 Cost estimation (D2.4iv)

The basis for the costs are defined and calculated in the Appendix. The basis for the cost comparison is shown in

Table 1. The Shell process described in Section 2 is used as the CCSM reference as this process has the most detailed and current cost data available.

Table 1. Comparison of transport and storage charges for CCSM and GCCS.

| | Cost/ tonne CO ₂ (£/tonne CO ₂) | | Power station (£M/year) | | Cement plant (£M/year) | |
|--------------------------|---|------|-------------------------|------|------------------------|------|
| | CAPEX | OPEX | CAPEX | OPEX | CAPEX | OPEX |
| CCSM | 20.5 | 20.5 | 6.8 | 6.9 | 11.5 | 11.5 |
| GCCS pipeline | 10.8 | 10.9 | 3.6 | 3.6 | 6.1 | 6.1 |
| GCCS Tanker | 15.5 | 15.6 | 5.1 | 5.2 | 8.7 | 8.7 |

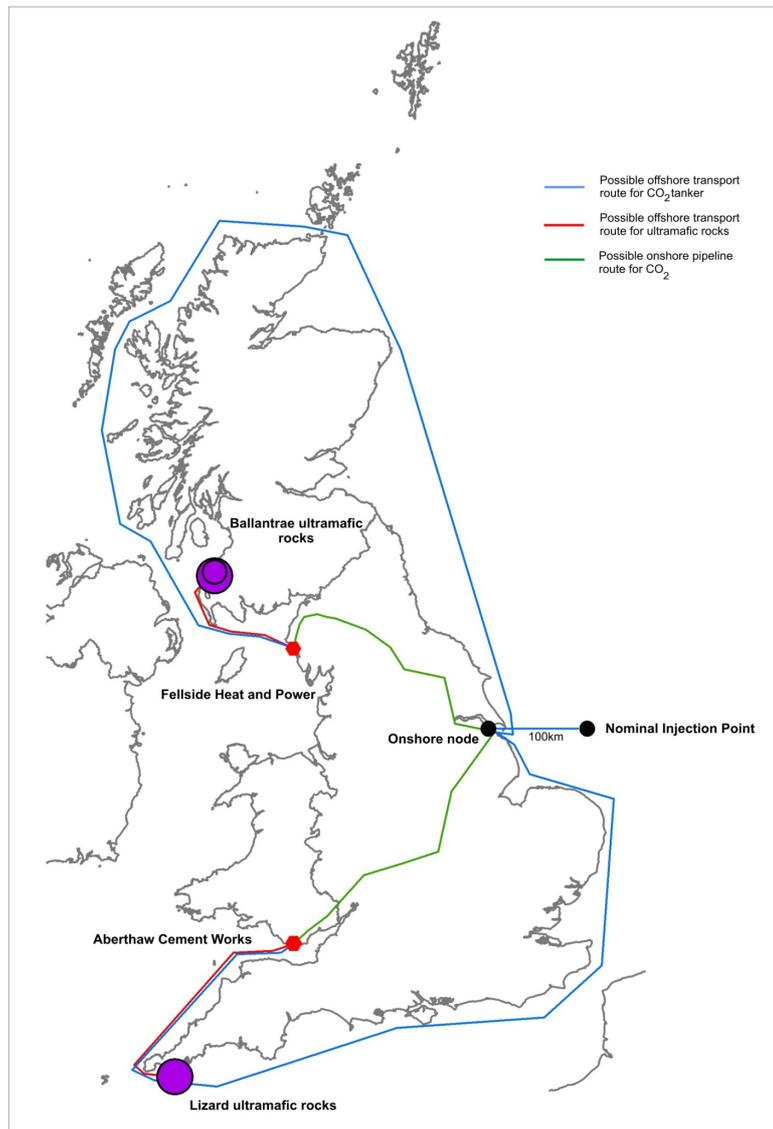


Figure 6. Map showing the locations of the sources of rock (purple circles), the emitters (red hexagons), the nominal routes of onshore pipelines and sea transport routes. For GCCS the CO₂ is taken to an onshore node terminal where it is transferred to an offshore pipeline to the injection site.

3.2.1 Assumptions

It is assumed that:

- The cost for transport does not include the capital costs of building ports, pipelines compression equipment, rock transport ships, or CO₂ tankers.
- Three tonnes of serpentine rock are required per tonne of CO₂ captured using CCSM (80% capture efficiency). This data is consistent with the Shell design where to capture 34kg/s of CO₂ about 107kg/s of raw material was required.
- Carbonation products are 33% heavier than the feed rock, and they will be returned to the source, but no landfill tax will be paid.
- No value is secured from sale of the carbonation products.
- Captured CO₂ will be liquefied before transport in a ship, and that CO₂ will be compressed before transport in the pipeline.

- No capital costs are included for transfer from the tankers to the pipeline, or on land pipelines to the offshore injection pipeline.
- The onshore node point is near Grimsby in north Lincolnshire and an offshore pipeline length of 100km reaches the offshore storage site.
- The onshore pipeline a cost of £6.30 per tonne CO₂ transported (ZEP, 2010). The costs are derived from a large-scale power plant with a capacity of 20Mt/year, and both CAPEX and OPEX were accounted. Pipeline costs consist mainly (~90%+) of CAPEX as reported in the ZEP report. However, the real costs for smaller CO₂ emitters might be much larger. Onshore pipeline was not considered feasible for distance >180km for such a low CO₂ volumes.
- The offshore pipeline cost is £7.40 per tonne (ZEP 2010). The costs are referred to a demonstration projects with a capacity of 2.5Mt/year, and both CAPEX and OPEX were considered.
- The cost for storage and monitoring estimated by McKinsey in 2008 is £8 per tonne CO₂ although this datum is currently unsubstantiated with long-term practical operating data.
- Onshore and offshore costs include capital cost of pipelines and compressor stations and operation costs. Storage costs include capital costs of an injection well and operation costs.

The assumptions outlined above are realistic, and the selected cost data are from reputable sources and sit within the range of sources reviewed. The team will seek improve data quality during the rest of the project.

3.2.2 CCSM and GCCS transport and storage cost comparison

The basis for the costs were defined and calculated in Appendix A1 and A2. Table 1 summarises the data.

The values used in

Table 1 depend significantly on the amount of CO₂ released by the two facilities and the assumptions listed in 3.2.1. GCCS cost is based on the ZEP report (2010), where cost is based on the same cost per tonne CO₂ as large-scale scenarios (5Mt CO₂/year). The level of accuracy in the CCSM and GCCS data is plus or minus 50%. The project team will seek to locate and generate figures with a higher level of certainty.

The overall transport and storage cost for CCSM comes to £41 per tonne of CO₂. This uses a combination of well-known and used components such as quarrying rock and sea transport of inert materials using bulk carriers. All the technology is tried and tested and requires no development. The main factor that drives the cost of CCSM is the utilisation of the magnesium cations by the carbonation process, and if this is improved, the amount of rock needed per tonne of CO₂ captured will fall and so will costs.

In summary it is concluded that for small and medium scale emitters:

- The cost of transport and storage for CCSM is broadly the same as the cost for GCCS, using the data and scenarios described here, and in both cases, it represents a relatively small part of the overall cost of CCS.
- There are several factors associated with the onshore pipeline that could make it more expensive, particularly distance from the emitter to disposal site and the treatment of gas required before transport.
- The costs for CCSM are based on well known, commonly used mineral industry processes.

- Costs for CCSM are virtually independent of the distance between rock source and emitter if both are at coastal locations and sea transport is used.
- The technical, health, safety, and environmental risks associated with transport and storage for CCSM are significantly less than GCCS. However, this is only one aspect of the risk profile for the processes, and a full systematic risk analysis will be required before any project moves forward.

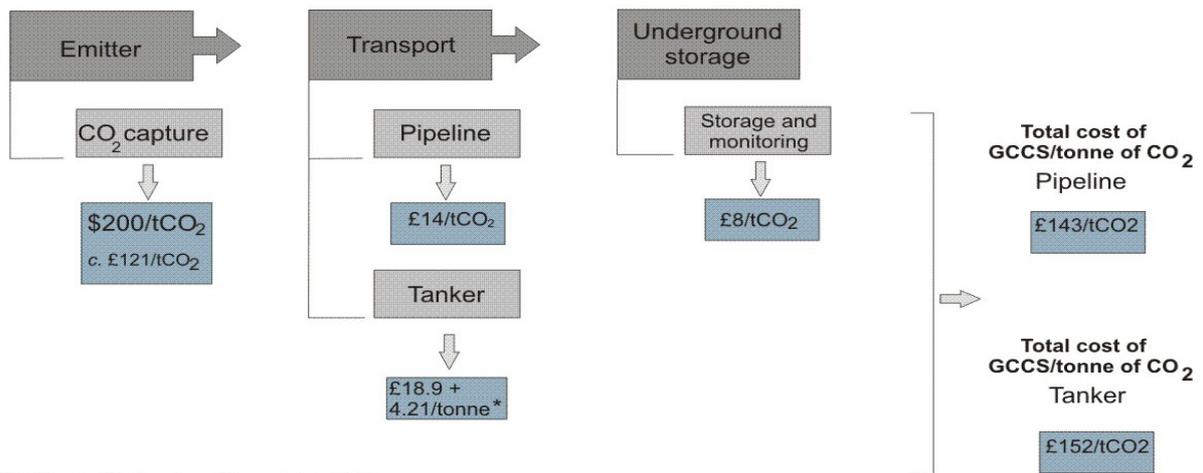
3.2.3 CCSM and GCCS capture costs

GCCS technology experience at various scales is very limited, and cost estimates and technology choices have a degree of uncertainty. Although much of the technology, such as amine scrubbing systems, has been operated for many years and is low-risk technology, there are still operational issues to resolve such as the purity of CO₂ required for large-scale high-pressure storage systems, whole system reliability and stability, well injection systems, and risk management of storage sites.

The consortium identified a number of different factors that explain the discrepancy between costs reported in different published reports. The most important are (IMC, 2008; McKinsey, 2008):

- Different time scale (e.g. costs provided by StatoilHydro are based on putting a GCCS plant in operation in 2018 while McKinsey is based on 2030).
- GCCS applied to steel and cement works and refineries has higher cost because CO₂ capture is not inherent in the design of these facilities. The collection of CO₂ from multiple flues, the contaminants in the flue gas streams and the small scale of some installations makes GCCS a challenging proposition.

For the above reasons, this assessment concludes that an average a GCCS capture cost for a small industrial emitter such as a CHP, cement plant or a gas-fuelled small power station would be around £121/tonne CO₂ avoided (including cost penalties for make-up production and incremental CO₂ emissions resulting from CO₂ capture). The avoided capture cost (£121/tonne CO₂) was estimated as the average of the range of £68/tonne to £174/tonne of CO₂ avoided reported for chemicals, fertilisers, refineries and small gas fuel operations (ICM, 2008). The main drivers for higher costs are smaller scale plants producing flue gases with different characteristics and CO₂ concentrations emitted in varying volumes and rates compared to a coal power plant (e.g. gas-fuelled power stations (5% CO₂), CHP plant used in crude-oil refineries (3.5% CO₂)). Adding transport (£23) and storage (£8) costs brings the total estimated cost of GCCS to £152/tonne of CO₂ captured. This is summarised in Figure 7.



IEA Energy Technology Essentials, 2006

* cost of liquefaction (4.21/tonne)

Figure 7. Summary of GCCS Costs.

The most comprehensive techno-economic assessment of the CCSM technology to date has been made by Shell. This indicates a cost of £180/tonne CO₂ with out financing charges and a cost of £327/tonne CO₂ for direct carbonation using flue gases instead of pure carbon dioxide including finance charges.

The cost of £327/tonne CO₂ is considered a realistic figure. Costs are summarised in Figure 8.

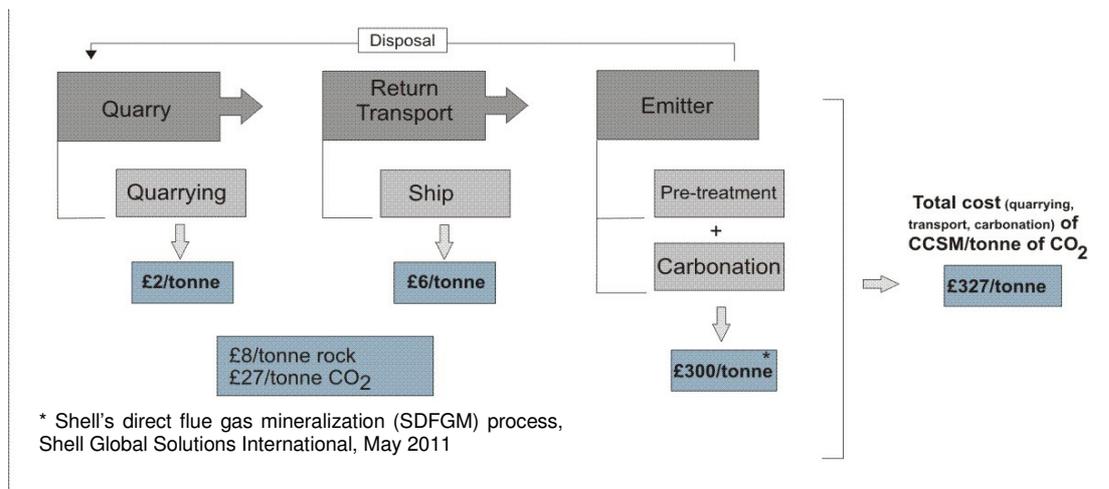


Figure 8. Summary of CCSM Costs.

3.2.4 Comparison of CCSM and GCCS costs

Overall, at the current level of knowledge CCSM costs are about double those of GCCS. This is about £330/t and £150/tonne CO₂, respectively, in the selected scenarios. The capture step is the most costly representing about 80% of the overall GCCS costs while transport and storage represent the remaining 20%. Similarly, the mineralisation step represents 90% of the CCSM costs. Therefore, the CCSM costs must be significantly reduced through further development and integration of different technologies in the process is to become an attractive investment.

3.2.5 Conclusions

The overall CCSM cost clearly indicates that the current high capital investments required are related with the mineralisation stage and are only marginally affected by transport and mining. Therefore, the focus of future experimental work needs to be focused on decreasing the capital cost and operating cost of the most capital and energy intensive components of mineralisation units. In order to accomplish this, experimental work will be focused on developing a clear and fundamental understanding of the reactions and process steps that could have most effect on the cost of CCSM. The following sections outline the main challenges and the project teams proposed activities to address each of them.

4 Common Requirements for Mineralisation Processes

Plants that could form effective carbon capture with mineralisation will require the following characteristics. They will have to:

- Handle large volumes of gas
 - This could be a low CO₂ concentration flue gas or be a process that includes CO₂ separation at an early stage in the process;
- Handle large volumes of solids
 - In the Shell base case the mineralisation plant for a 300MW power station will be required to feed and process between 350 tonnes and 400 tonnes of rock each hour;
- Handle the rock in a way that creates large surface area.
 - This requires the sufficient grinding to create rock particles with a diameter of less than 50µm;
- Activate the mineral through thermal or chemical processes to make it receptive to react with the carbon dioxide;
- Be large scale process plants:
 - Current indications are that they will have to be larger than the plants from which the carbon dioxide is produced;
- Increase kinetic and/or mass transfer rates to increase rate of reaction and hence reduce residence time and plant size.

All of these issues are significant challenges that the development of all CCS mineralisation processes need to address before they can become effective in the capture and storage of CO₂.

4.1 Technology Readiness Levels

As part of this overview the team has looked at the current technology readiness level (TRLs) of the various technology development options that could be pursued by the consortium. The following table (Table 2) summarises the description of the TRLs as originally developed.

Table 2. Technology Readiness Levels.³

| | Technology Readiness Level | Description |
|---|--|--|
| 1 | Basic principles observed and reported | Lowest level of technology readiness. Scientific research begins to be translated into applied research and development. Example might include paper studies of a technology's basic properties. |
| 2 | Technology concept and/or application formulated | Invention begins. Once basic principles are observed, practical applications can be invented. The application is speculative and there is no proof or detailed analysis to support the assumption. Examples are still limited to paper studies. |
| 3 | Analytical and experimental critical function and/or characteristic proof of concept | Active research and development is initiated. This includes analytical studies and laboratory studies to physically validate analytical predictions of separate elements of the technology. Examples include components that are not yet integrated or representative. |

³ Technology Readiness Levels in the Department of Defense (DoD), (Source: DoD (2006), *Defense Acquisition Guidebook*)

| | | |
|---|---|---|
| 4 | Component and/or breadboard validation in laboratory environment | Basic technological components are integrated to establish that the pieces will work together. This is "low fidelity" compared to the eventual system. Examples include integration of 'ad hoc' hardware in a laboratory. |
| 5 | Component and/or breadboard validation in relevant environment | Fidelity of breadboard technology increases significantly. The basic technological components are integrated with reasonably realistic supporting elements so that the technology can be tested in a simulated environment. Examples include 'high fidelity' laboratory integration of components. |
| 6 | System/subsystem model or prototype demonstration in a relevant environment | Representative model or prototype system, which is well beyond the breadboard tested for TRL 5, is tested in a relevant environment. Represents a major step up in a technology's demonstrated readiness. Examples include testing a prototype in a high fidelity laboratory environment or in simulated operational environment. |
| 7 | System prototype demonstration in an operational environment | Prototype near or at planned operational system. Represents a major step up from TRL 6, requiring the demonstration of an actual system prototype in an operational environment, such as in an aircraft, vehicle or space. Examples include testing the prototype in a test bed aircraft. |
| 8 | Actual system completed and 'flight qualified' through test and demonstration | Technology has been proven to work in its final form and under expected conditions. In almost all cases, this TRL represents the end of true system development. Examples include developmental test and evaluation of the system in its intended weapon system to determine if it meets design specifications. |
| 9 | Actual system 'flight proven' through successful mission operations | Actual application of the technology in its final form and under mission conditions, such as those encountered in operational test and evaluation. In almost all cases, this is the end of the last "bug fixing" aspects of true system development. Examples include using the system under operational mission conditions. |

It is clear from the work undertaken so far that the chemical reactions behind the potential processes are well researched, and it is clear that they have the potential to be used as carbon capture media. However, this fundamental reaction understanding has not yet been converted into a viable industrial scale CCSM process. There are two groups of process types that the next stage of the process development work will develop further. These are described in detail in the full Stage 1 report⁴.

The direct and indirect aqueous processes that use additives were identified as the most promising processes to be investigated for a better understanding of the process fundamentals (reaction rate and kinetics) towards the reduction of the CCSM CAPEX and OPEX. In particular, processes that use Na/ammonium-based additives were selected for further investigations. These processes have been developed to a range between TRL 2 and 4.

Overall carbon capture through mineralisation is still an early stage technology that is far from scale up, and it is likely to require a concerted investment over a period of more than 10 years to develop it to a stage where it will be able to attract significant investment. However, much of the necessary technology is well understood, and an outline of the potential development route is discussed below.

4.2 The Unit Operations of the Processes and the Targets For Development

As stated above some elements of the integrated mineralisation system have been in regular use for very many years, while others are novel and require significant development.

In looking at process improvement activities a high level assessment of costs has been made. The costs split for the Shell process are as shown on Table 3 below.

⁴ Full Stage 1 report for the ETI Mineralisation Project.

Table 3. Comparison of Pre-treatment and Carbonisation Costs for CCSM.

| CCSM costs | £/tonne CO₂ |
|-------------------|-------------------------------|
| Pre-treatment | 100-150 |
| Carbonation | 150-200 |
| Total | 300 |

4.2.1 Mining and Quarrying Technologies

Rock winning and comminution is well-understood technology that, in some cases, has been in use since the early industrial revolution. However, a wide range of equipment is used in the carbon capture process. Technology includes rock winning and crushing to < 200mm for rail transport followed by further crushing to 25mm for introduction into the Shell process. The costs of this upstream treatment do not appear in the Shell capital estimates¹. The logistics costs have been studied in section 3, but the upstream crushing and processes cost require further analysis. There is then a requirement for ball milling to reduce the serpentine to the 50 microns required for the process. This fine wet and dry grinding is included in the Shell process at a cost of £163 million installed. However, there are some recent developments in the comminution market that may offer process improvement opportunities.

4.2.2 Thermal Activation

The Shell process uses a thermal activation to increase the reactivity of the serpentine minerals at an installed capital cost of £150 million. As a result of the high temperatures required, waste heat from the process cannot be used for this purpose, and there is a large parasitic heat load that is supplied by imported natural gas. Operating costs for thermal activation are between £100 and £150/tonne of CO₂ captured. Eliminating or reducing the cost of this step would have a significant impact on both capital and operating costs of the plant. The activation of the mineral prior to use in the process is one of the major targets for process improvement.

4.2.3 Carbon Dioxide Absorption Reactor Technology

Another major target for process development lies in the development and proving of the complex absorption reactors required by all of the chosen processes. In the Shell process, this step incurs a capital cost of £180 million. The processes need to be improved and demonstrated for all potential reaction routes. This technology is so far unproven above pilot scale demonstration. As the SGS process utilises a thermal activation step, it is believed that this can be improved or possibly eliminated in a modified process route. In addition, there are opportunities for improvements in all process steps that have the potential to reduce capital and operating costs. Further experimental testing and development is required to define optimum reaction conditions, and this work will form the major part of the next phase of development work on Stage 2b.

4.2.4 Mineral Precipitation

The mineral precipitators (including the thickeners, liquid precipitators and slurry precipitators) are also very significant items of process equipment with a capital cost of £225 million for the Shell process. Any improvements in these units will also have a significant impact on the overall plant size and capital cost.

4.2.5 Process Design

The final part of the process requiring further work is the basic process design. This will be undertaken through a combination of routes including the existing Shell Aspen model, but also including outline designs for all processes. It is hoped that improvements in the design of the

process plant elements will reduce size and scale of the surrounding equipment, and the proposed development activities are targeted at delivering these process improvements. It is likely that this type of approach will reduce process scale and cost.

4.3 Technology Challenges

There are a number of significant technology challenges that the project team is focusing its efforts on for the remainder of the programme. These are:

- Reducing the capital plant through process redesign, changes to the operational characteristics and the development of additional reactor technology;
- Efforts to reduce capital will also seek to reduce operational energy consumption to reduce the parasitic load of operation and as a result raise the amount of carbon dioxide that can be captured;
- The plant designs to date are physically very large and work will be undertaken to define options to reduce plant size. Again this is related to operational efficiency and is strongly linked to overall plant capital;
- The development of enhanced reactor technology will be a long journey from the current lab and semi-pilot trials to a fully-fledged carbon capture process. To aid this development the project team are undertaking additional work in the following areas:
 - Additional experiments to assess changes in reaction performance between different UK serpentine rock types,
 - Increasing the rate of Mg extraction from serpentine,
 - Increase rate of CO₂ uptake in the carbonation reaction,
 - Redesign to decrease the size/volume of process reactors
- To use this work to clarify the effect of changing rock, flue gas and contaminants on reaction kinetics and reactor design;

4.4 Conclusions and Technology Development Opportunities

We have looked at two coastal sites for implementation of CCSM in UK: the Fellside Heat and Power plant that emits 0.56Mtpa CO₂, and Aberthaw Cement plant in Glamorgan that emits 0.33Mtpa CO₂. The CCSM cost for the selected sites was estimated and compared to the cost of alternative options such as GCCS using (a) pipelines and (b) tanker.

We have found that for the selected scenarios:

- The total cost of CCSM is almost twice higher than GCCSM:
 - CCSM, £327/tonne CO₂ > GCCS tanker, £152/tonne CO₂ > GCCS pipeline, £143/tonne CO₂.
- The ratio of CAPEX:OPEX is approximately 50:50 for all scenarios, and the cost of transport and storage for CCSM and GCCS is in the order:
 - CCSM, £41/tonne CO₂ > GCCS tanker, £31.1/tonne CO₂ > GCCS pipeline, £21.7/tonne CO₂.
- The cost of CCSM is virtually independent of the distance between mineral deposit and emitter when both are at coastal locations and ship transportation is used.

The cost of CO₂ capture for GCCS reported in the literature is the most uncertain parameter and varies from £45 to £170 per tonne CO₂ and accounts up to 80% of total GCCS cost; similarly, the cost of carbonation step for CCSM accounts for 90% of total CCSM cost. Therefore, CO₂ carbonation step presents the main opportunity for the overall CCSM cost reduction.

This can be achieved through selection, design, and optimisation of large-scale chemical carbonation reactors based on in-depth understanding of process fundamentals and related process control parameters.

Table 4 summarises the key figures from this and the TEA-1 report.

Table 4. Summary of CCSM costs.

| | Cost | Comment |
|--|------------------|---|
| Cost of capital equipment for a Shell technology CCSM Plant | £107m | Outlined in section 2. |
| Installed capital cost of CCSM plant using Shell Technology | £720m | Installation factor is at the high end of the range, but is realistic given the complexity of the plant |
| Percentage of CO ₂ emitted that is avoided | 50% | There is a large parasitic load due the thermal activation process |
| Total cost of capture per tonne of CO ₂ avoided for CCSM | £305/t | Assumes cost of capital is 10% |
| Approximate cost of pre-treatment and activation per tonne of CO ₂ avoided for CCSM | £100/t to £150/t | This figure is linked to the pre-treatment process |
| Approximate cost of carbonation per tonne of CO ₂ avoided for CCSM | £150/t to £200/t | |
| Total cost of capture per tonne of tonne of CO ₂ avoided for GCCS | £140/t to £150/t | |
| Cost of carbon avoidance per kWh using current CCSM technology | £1.45 kWh | |

Currently, mineralisation processes require plants that are physically too large to be viable and Table 4 leads to the conclusion that they are too expensive to justify investment. However, there are indications in the work conducted to date that improvements in cost and design can be made that could reduce these costs by a factor of two.

The consortium has outlined these detailed proposals in its Stage 2b Development Proposal. In summary the proposed next steps are:

- Identify plant elements that are common to all processes and seek to design lower cost components or locate lower cost technology;
- Carry out lab and semi-pilot experiments to Improve reagent reactivity and reactor design looking at the process types identified by the team;
- Work to identify and develop cheaper process design options;
- Redesign the process to reduce parasitic energy consumptions;
- Develop more efficient reactor designs

Once the next phase of experimental and design work is completed the consortium will have a much clearer idea of the true capital and operating requirements based on technology that is low cost but can do the job. It will also have a clearer idea of the process development opportunities that create more compact and lower energy processes.

The lower capital, smaller footprint plant should have more favourable economics and the scale of the improvement can be assessed.

At the end of Stage 2b a further Go/No Go review should be undertaken. If the work is successful and an economically acceptable way forward can be identified work should then start to develop an Aspen simulation of the continuous pilot scale process to enable a more detailed techno-economic analysis to be performed.

5 CCSM Project Plan Forward

5.1 CCSM Technology Quantitative Comparison (D2.4v)

As it was discussed in Stage Gate 1 Report, the deployment of mineral carbonation is dependent on the development of low energy-consumption processes with high mineral reaction conversion and fast reaction kinetics. To date demonstration has largely used pure CO₂ or simulated flue gases. Twenty-five CCSM process routes were described in Chapter 14 of the Stage Gate 1 Report. These were ranked against the 5 major criteria described in Chapter 10. This ranking was revised using a Pugh method where the CCSM process developed by ARC/NETL was used as Datum. In addition, the criteria were weighted as follows: Energy need: 5, Mineral requirement: 3, Valuable products: 3, Apparent simplicity: 1, Apparent reliability: 3. The summary is shown in Table 5.

Table 5. Pugh Analysis Results Using ARC-Serpentine CCSM as Datum.

| Process/Criteria | Energy need | Mineral req. | Valuable products | Apparent simplicity | Apparent reliability | Initial Rating | Pugh ranking |
|--------------------|-------------|--------------|-------------------|---------------------|----------------------|----------------|--------------|
| Schiller | A | B | D | C | D | 16 | 4 |
| Huijgen | B | B | C | B | C | 18 | 2 |
| Brent | B | C | B | A | B | 20 | 0 |
| ARC- serpentine | B | C | D | A | A | 19 | 0 |
| Hunwick | B | C | B | B | C | 18 | -1 |
| ARC- olivine | E | B | D | A | A | 17 | -2 |
| Maroto-Valer | A | D | D | C | B | 16 | -2 |
| Shell- Olivine | - | A | C | B | B | 16 | -3 |
| Baclocchi (wet) | - | B | C | C | C | 13 | -3 |
| Baclocchi (dry) | - | B | C | C | B | 14 | -3 |
| Caterpillar | C | B | C | B | B | 18 | -3 |
| Åbo Akademi | B | D | B | B | B | 18 | -4 |
| Kodama | B | C | D | C | D | 14 | -4 |
| Shell- serpentine | - | - | C | A | A | 13 | -5 |
| Baldyga | - | A | D | - | C | 10 | -6 |
| N'ham process | - | - | C | B | B | 11 | -9 |
| Kwak | E | E | C | D | D | 9 | -9 |
| Calera Corp. | - | - | - | - | A | 5 | -9 |
| Park | - | C | D | B | C | 12 | -9 |
| Lin <i>et al.</i> | E | E | D | C | C | 10 | -12 |
| Krevor | - | C | D | D | C | 10 | -12 |
| DaCosta | - | - | - | - | B | 4 | -12 |
| Rau & Caldeira | - | - | - | - | - | 4 | -15 |
| Munz <i>et al.</i> | - | - | - | - | - | - | -15 |
| Reddy | E | - | C | C | D | 9 | -15 |
| Vandor | - | - | - | - | B | 4 | -15 |
| Jones | - | - | - | - | A | 5 | -15 |

Table 5 summarises the analysis. It shows that only the group in the box is the same or better than the base case which was the ARC CCSM. The top groups have the following characteristics. They are:

- aqueous
- single-step or multi-step
- with additives (Na-based or NH₄-based) or w/o additives
- improved ARC process by utilisation of waste heat (Brent)

Overall, no single technology demonstrated ALL of the following essential criteria required for the large-scale implementation:

- CO₂ capture efficiency 80% or better
- Mineral utilisation 80% or better
- Low energy consumption in terms of operating T, P and duration
- Demonstrated robustness to flue gases (rather than pure CO₂ stream)
- Low chemical additive consumption and recyclability

The technologies listed in the table can be grouped as follows:

- Group 1: Direct gas-solid reactions (DaCosta, Reddy, Baiocchi (dry))
- Group 2: Direct, aqueous reactions without additives (Kwak, Munz, Baciocchi (wet))
- Group 3: Direct aqueous reactions with additives (Brent: (Na-based) plus the ARC-olivine/serpentine, Calera, Caterpillar, Jones, Krevor: (organic solvent)
- Group 4: Indirect gas-solid reactions (Abo Akademi, Lin: (strong acid and base))
- Group 5: Indirect aqueous reactions with additives (Schiller: (ammonium-based). University of Nottingham, Kodama, Hunwick, Shell (Na-based) and Park: (organic solvent)

CCSM technologies can be grouped around the attractive features of individual methods within the same group and/or between the groups, shown in Figure 9. There are two basic types: direct and indirect carbonation and then, further divided into 5 final sub-groups.

The **direct gas-solid process** (group 1) requires temperatures between 150-500°C and fine grinding of minerals (5-75µm); the CO₂ capture efficiency of 75% was demonstrated with highly reactive waste feedstock (see Baciocchi and Reddy) but <40% efficiency was achieved with minerals.

Indirect gas-solid carbonation process (group 4) includes the extraction of Mg(OH)₂ from serpentine with additives (ammonium sulphate for the Akademi process). This approach demonstrates better mineral utilisation, but requires a temperature between 450-550°C for the carbonation step; maximum CO₂ capture efficiency is below 60%. Gas diffusion through the carbonation product layer is the key factor limiting the efficiency of the overall process.

Overall, gas-solid pathways (both direct and indirect, groups 1 & 4) do not show advantages over alternative methods. Contrary, it is well documented in the literature that the presence of water considerably enhances the reaction rate in the carbonation process; therefore it is recommended that gas-solid routes are not considered further in this project.

The **direct single step aqueous carbonation** (group 2) requires high pressure between 20-150 bar and temperatures between 150-500°C with a particle size between 10-37µm and a reaction

time of up to 20 hr. CO₂ capture of maximum 50% was achieved due to poor dissolution of minerals in water as well as limited CO₂ solubility at high temperatures.

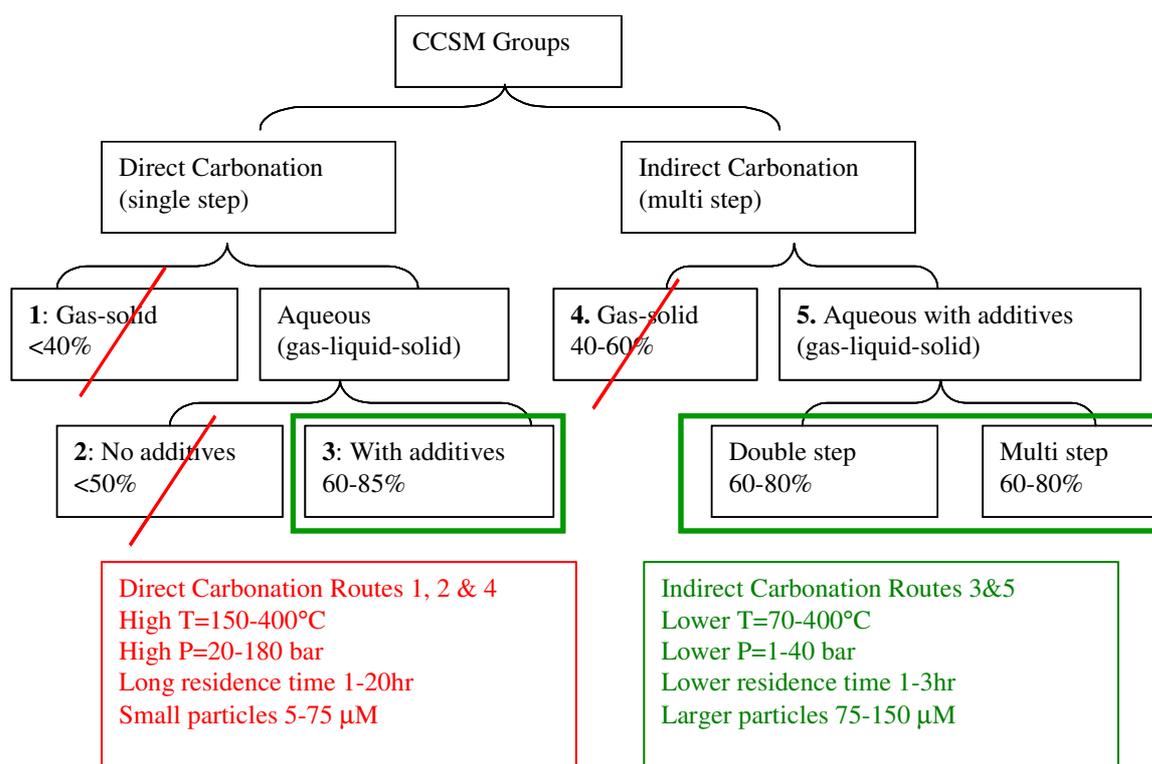


Figure 9. Different CCSM pathways (modified after Delgado, 2010). The numbers 1 to 5 indicate groups of similar CCSM technologies. Percent value shows the CO₂ capture efficiency calculated as $(CO_2^{in} - CO_2^{out}) / CO_2^{in}$.

Direct single step aqueous carbonation with additives (group 3) showed significant improvements over group 2 with CO₂ capture efficiency up to 85% with pure CO₂ stream (ARC process) and 30% with flue gas (Shell process) at temperatures between 140-150°C. Meanwhile, the regeneration and recyclability of additives (NaOH, NaHCO₃) will have to be addressed. In the Caterpillar process, the alkaline waste streams are used as additives. CO₂ capture efficiency was ~25%, but this is due to a short residence time in a spray dry absorber and utilisation of simulated flue gas.

The **2-step and multistep aqueous processes with additives** (group 5) are able to reach high carbonation efficiency using mild process conditions and short residence time as a result of faster reaction kinetic in presence of additives. The pros and cons are summarised below.

| Group 5 Pros | Group 5 Cons |
|--|---|
| <ul style="list-style-type: none"> - Uses common chemical agents used in mineral processing so does not require any specialised equipment. - Operates at < 20 bar, < 150°C for 0.2-2 hours. - CO₂ capture efficiency 60-80%. | <ul style="list-style-type: none"> - Multiple heater/coolers, pumps and compressors. - Material losses are unavoidable, including CO₂ slip. - Additive regeneration has not been achieved so far. |

A wide range of additives for groups 3 & 5 allows enhancement of the carbonation efficiency. The use of strong acids or strong bases is limited due to corrosion and toxicity⁵. The use of catalytic enzymes such as carbonic anhydrase is unlikely to be effective due to instability and low activity above 170°C⁶.

The recyclability, availability and accessibility of additives represent a challenge for aqueous CCSM technologies in groups 3 and 5. Electrolytes such as NaCl, NaHCO₃ and ammonium salts are recyclable so are attractive. These additives are key components in recent CCSM developments (ARC, Shell, Abo, UoN). In addition, CO₂ scrubbing with ammonia is well-established technology and is similar in operation to amine scrubbing systems for GCCS. Ammonia-based compounds are the most promising additives for the group 3 & 5 CCSM technologies.

The ARC feasibility study indicates sequestration costs of £156/tonne with the carbonation reactor capital cost representing 2/3 of the overall costs⁷. This study highlights ways to de-risk the carbonation technology and to decrease the technology costs. The above assessment confirms the output of the SG1 report and indicates that the reactor kinetics are a key aspect in lowering costs (Figure 10), where low residence time and low operation pressure would drastically reduce the costs of the ARC process. They conclude that accelerating the kinetics and decreasing residence time could decrease the sequestration costs from £156/tonne to £68/tonne. Similarly, they show that decreasing the pressure from 160 bar to 17 bar would reduce the capital costs from £740M to ~ £120M. In turn, this would result in a CO₂ sequestration cost of £22/tonne. This shows that there are significant opportunities for capital improvement through redesign and energy integration combined with the extraction of valuable products such as iron oxide from the mineral output (see Brent).

The Shell process uses the same additives (NaHCO₃), pre-treatments (thermal treatment and very fine grinding to 20µm) and reaction time (2hr) as the ARC process, but operates at lower temperature (140°C) and pressure (30 bar). It can also operate on flue gas, although the capture efficiency drops below 50%. The cost including capital charges is around £320/tonne CO₂ avoided without product drying. However, no optimisation experiments have been carried out. Therefore, the optimisation of the process conditions (liquid-to-solid ratio, T, P, time, particle size) used in the Shell process together with the integration of other process units could sensibly decrease the overall CCSM technology costs for this process.

The full economics of the group 5 processes based on ammonia compounds (Kodama, Åbo Akademi, UoN) are not currently known. However, NH₄-based processes have at least two significant advantages over Na-based ones: (1) they allow CO₂ capture from flue gas, avoiding additional separation and compression stages and (2) mineral dissolution can be performed with larger particle sizes (from 75 to 150µm), resulting in less energy requirements for mineral grinding. Furthermore, various integration options need to be tested (e.g. less severe operating conditions such as operating at low pressures using higher solid/liquid ratios and integration of separated units such as activation and CO₂ absorption) to lower capital and operational costs.

⁵ Pundsack, 1967; Lackner et al., 1995; Fouda et al., 1996; Park et al., 2003; Maroto-Valer et al., 2005; Blencoe et al., 2003; Wendt et al., 1998; Lackner, 2002.

⁶ Brent, 2010; Huijgen and Comans, 2005.

⁷ C. White, 2003.

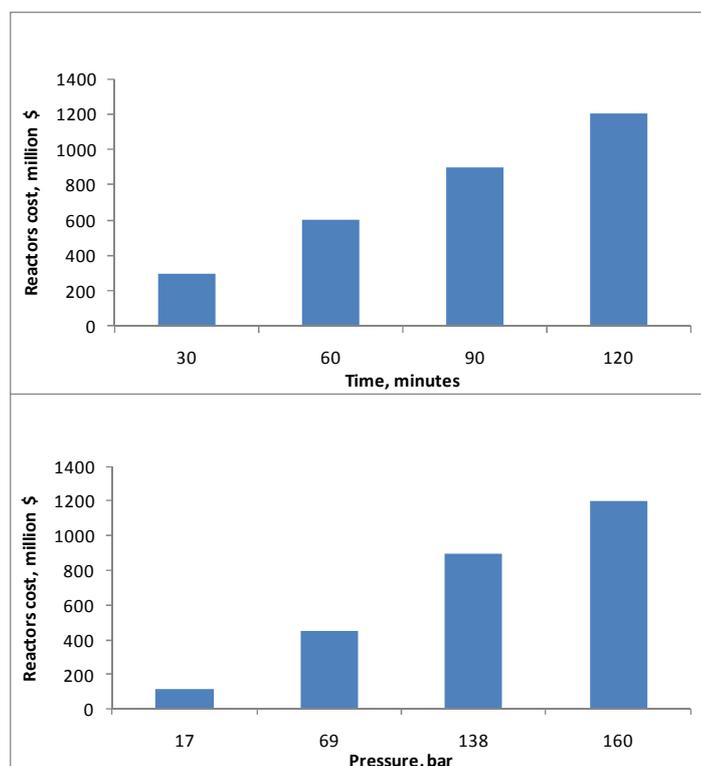


Figure 10. Effect of time and pressure on the CCSM capital costs where 160 bar and 120 minutes were the assessed conditions (modified after White, 2003⁷). Sensitivity based on 95 reactor vessel of 378.5m³ able to treat 24,000 tonnes CO₂ per day.

5.1.1 Technology Comparison Conclusions

The comparative analysis of CCSM technologies shows that:

- aqueous CCSM are preferred (all top 5 CCSM in Pugh table are aqueous);
- additive enhanced aqueous CCSM are preferred (group 3 and group 5);
- aqueous additive enhanced multi-step processes (group 5) might be more promising but additional experimentation and analysis should be done for final recommendation (both capital and operational);
- Na and NH₃ type additives are most promising and final selection requires additional experimentation and analysis toward opportunity for cost reduction (both capital and operational).

Also, the study of the different technologies indicates that the acceleration of the dissolution and carbonation kinetics and the reduction of the residence time are the most important aspects towards the reduction of the CO₂ sequestration costs. To improve the reaction kinetics the experimental plan needs to be focused on increasing the rate of Mg extraction and increase the rate of CO₂ reaction.

5.2 CCSM Experimental Plan

The experimental process flow, roles and responsibilities are shown below (Figure 11).

The main goal of the experimental work is to identify and quantify the key factors that impact on CCSM costs and hence identify the specific opportunities and areas for cost reduction. As

indicated earlier in this report, the accelerated kinetics and the residence time reduction are the most important aspects that need to be addressed in the experimental work. This needs to be coupled with an assessment of the reactivity of UK resources. This knowledge will provide data for the follow up detailed TEA of the selected technology in application to two selected “mineral site / emitter” scenarios.

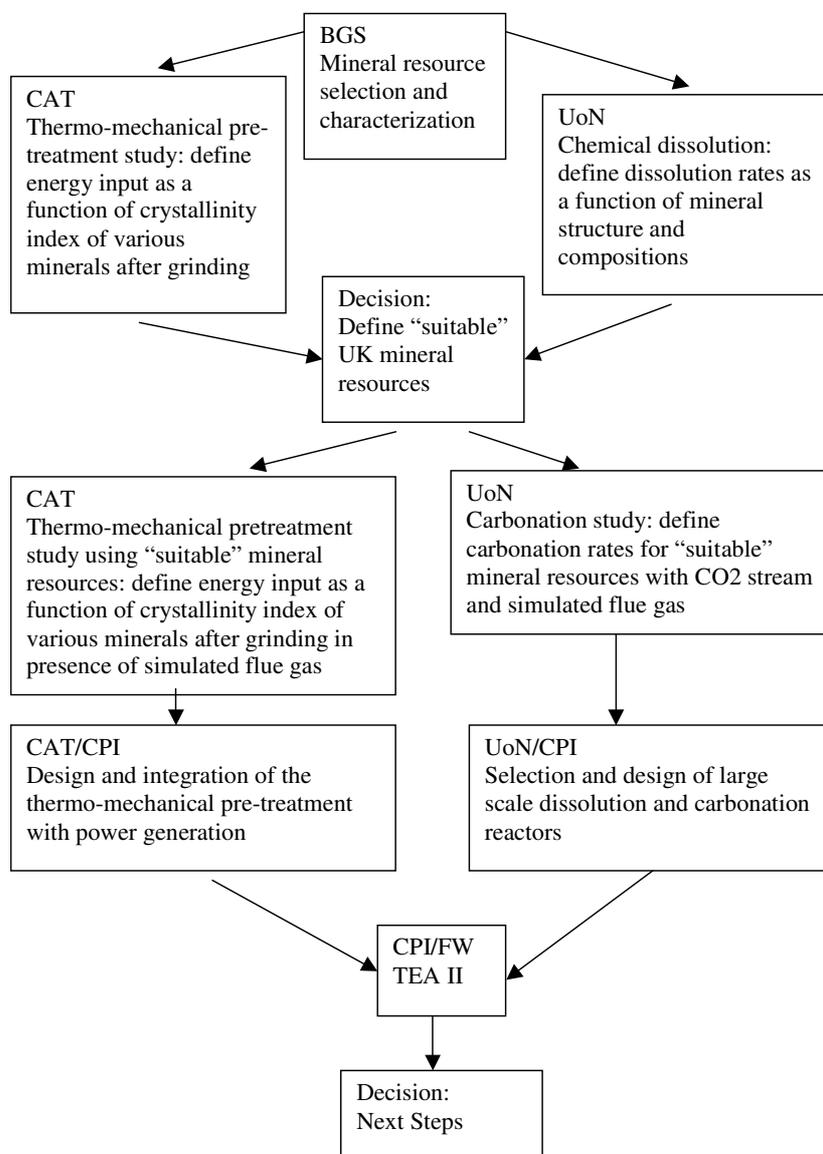


Figure 11. Experimental plan process flow.

5.2.1 Rock resources (D2.4ii)

Chapter 2 of the Stage Gate 1 report describes the diversity of ultramafic rocks and highlighted the fact that only a very few simple end member compositions have actually been investigated in mineralisation experiments. Estimates of potential resources of ultramafic rock feed material, in the published literature and those in this report, assumes that all ultramafic rocks have broadly similar behaviour during the mineralisation process. This assumption is necessary as there is little or no evidence to show whether it is right or wrong.

The limited amount of data is confused and contradictory. A clear shortcoming of most data is that the geological knowledge of the starting materials and accurate characterisation is lacking. The technologists did not really know what they were putting into their experiments and this undoubtedly partly explains the confusing results.

There is a need for a series of consistent mineralisation experiments on a range of well characterised rock compositions to establish the variation in reactivity of the various mineral proportions. These could be natural rocks or mixtures of separated mineral concentrates.

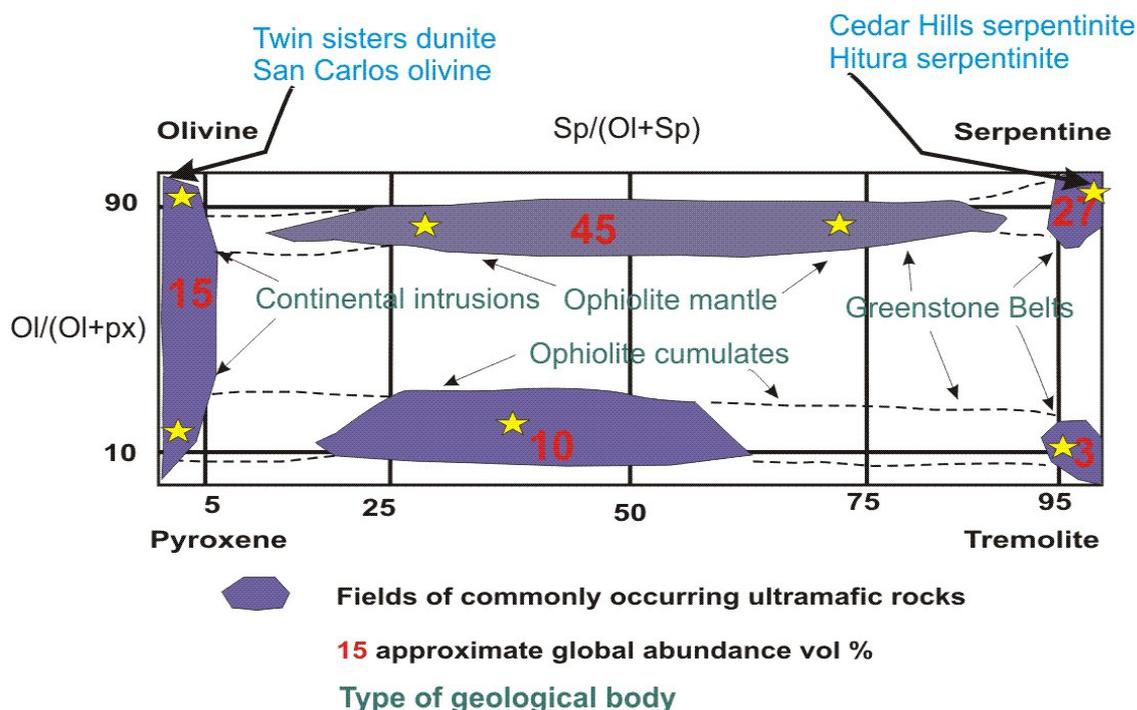


Figure 12. Yellow stars superimposed on the diagram for composition variation show the compositions of samples suggested for mineralisation experiments. Copyright BGS, NERC

Figure 12 shows the composition of the rocks or mineral mixtures required to define the reactivity of the main mineral components of ultramafic rocks and common natural mixtures. The pure olivine and serpentinite samples will help with comparison with previous studies while the others will help to define the reactivity of pyroxene and amphibole as well as the common rocks composed of mixtures of minerals. In addition to these, it is suggested that a chlorite-rich rock or mineral concentrate should be tested as chlorite is a common accessory mineral. It often forms around 5% of rocks but can be as much as 25%. However it is rarely the main component.

Once the basic reactivity of the range of minerals and rock types has been established it is important to determine how the various physical and chemical properties of the minerals such as particle size and composition affect the reactivity. The previous experiments should help with the interpretation of previous laboratory studies but the poor characterisation of materials used might make this difficult. A series of experiments to improve the data on the effect of these parameters is necessary and will form part of the WP2 activities.

4 samples from all the A-rated sites will be tested to generate real data. Six 10kg samples will be collected from each of the locations for testing. One large 50-100kg sample will be taken from Ballantrae to be used as a typical UK serpentinite for laboratory investigations in WP2.

The rocks used for experiments and end-products will be characterised by a combination of optical microscopy and SEM to characterise the minerals present, their crystal size and interrelationships of the different minerals, XRF to determine the bulk chemical composition and quantitative XRD to measure the proportions of minerals present in a bulk sample and the crystal structure of the component minerals.

During Stage 2B BGS will:

- (a) Carry out a detailed assessment of mineral accessibility to ascertain which bodies with potential in terms of size, do not have drawbacks that make them unsuitable for exploitation and therefore not a realistic resource.
- (b) Perform the assessment of existing samples to refine data on the mineralogy and composition of potential resources.
- (c) Identify types of data needed to adequately classify rock bodies and refine the “suitable resource” definition.
- (d) Perform laboratory investigation of additional samples from the UK; collect data on the characteristics of mineral deposits of interest to refine the assessment of their suitability as a feed material.
- (e) Assess existing data for global resources: this task aims to classify the main types of occurrence of ultramafic rocks and serpentinite in to geologically related groups and to show the distribution of the areas where these associations are found on a world map.

5.2.2 Thermo-mechanical activation of minerals (D2.4vii).

Chapter 11 of the SG1 report described the mechanical and thermal activation of minerals. It was shown that mineral pre-treatment cost and related energy usage is highly dependent on type of mineral (crystal structure, composition, toughness) and grinding conditions (type of mill, speed, wet vs. dry, etc.). This step of the CCSM process presents an opportunity for significant cost reduction through integration of heat, mass and mineral activation. Although integrated thermo-mechanical activation processes are known, its application to CCSM was not studied. During Stage 2b CAT will lead the development of thermo-mechanical activation and complete the following tasks:

- (a) Estimate an energy usage for mechanical activation of minerals as a function of feedstock variation (chemical composition, particle size, etc.). In order to derive this functional correlation, the energy input will be calculated (Bond index, grinding speed, duration, etc.) and the mineral crystallinity index, after grinding, measured (the ratio of characteristic XRD peak intensity before and after mechanical activation, I_0/I_f). In addition, surface area measurements of the activated mineral samples will be performed. The “suitable” UK mineral resources will be selected
- (b) Estimate an energy usage for heat-treat activation of minerals as a function of heat input. The heat energy input will be calculated (temperature, duration, mineral mass, heat capacity, etc.) and mineral crystallinity index measured. In addition, surface area measurement will be performed. The functional correlation will be created for selected “suitable” UK minerals and built into TEA-II cost and energy usage evaluation.
- (c) It is known that flue gas is acidic due to the presence of sulphur and nitrogen oxides. This environment may accelerate thermo-mechanical disintegration of minerals and decrease energy input. A set of thermo-mechanical activation experiments will be performed in a simulated flue gas. The activation energy consumption will be assessed against feedstock variability, temperature, grinding speed, duration and the mineral crystallinity will be measured. Surface area measurement will be performed to assess the available reaction surface. The functional correlation can be created for several representative minerals and built into TEA-II cost and energy penalty evaluation.
- (d) The energy consumption models for cases (b)-(c) will be compared. A set of optimum operating conditions for thermo-mechanical activation will be proposed to reduce cost and energy input. The solution will be validated in Stage 3.
- (e) The following equipment will be used: (i) high energy attrition mill with various grinding atmospheric conditions as well as cooling/heating options; (ii) high temperature furnaces

up to 1100°C with controllable heating/cooling rates and various gas atmospheres; (iii) Bruker D8 XRD for crystallinity index measurements; (iv) Micromeritic BET ASAP 2020 for surface area measurement of mineral samples; (v) solid state Fourier transform infra red measurement.

United States Department of Energy methods will be used.

5.2.3 Chemical activation (D2.4vii)

The University of Nottingham (UoN) will work on: (a) mineral activation by chemical leaching using Na and NH₄ additives and (b) the study of a multi-step aqueous carbonation process. Chemical dissolution experiments will be run on UK serpentinite rocks from 5.2.1 to compare the efficiency of the technique with typical US serpentine dissolution data from 5.2.2. Energy efficiency, thermal treatment and mechanical activation will be compared. The deliverable will be a clear data set for the effect of pre-treatment on UK resources.

During Stage 2B the UoN will:

- (a) Run chemical activation experiments and measure how dissolution rate of real rock samples is linked to chemical and mineralogical characteristics. This will identify the best scenarios where the CCSM could be deployed and closest rock resources.
- (b) Estimate the energy consumption for dissolution as function of feedstock variation (identified in *a*) in a controlled environment (additive type, additive concentration, pH, fixed T, P, duration, L/S ratio, etc.). These experiments will help the definition of “suitable UK mineral resources”.
- (c) Select mineral samples to measure the dissolution rates and energy consumption as function of additive type, additive concentration, pH, particle size, T, P, L/S ratio, and duration. These experiments will establish the conditions that maximise dissolution and minimise energy requirements.
- (d) Define the most promising Na and NH₄-based additives and compare that with data from ARC and Shell to make a final additive selection. A series of ammonium-based chemical dissolution experiments will be run as function of concentration (or L/S ratio) and pH.
- (e) The following equipment is available: X-Ray Diffraction (XRD), Inductively Coupled Plasma-Mass Spectrometry (ICP-MS), Thermo Gravimetric Mass Spectrometry (TGA-MS), and XRF. The equipment will be used for the qualitative and quantitative analysis of the solution with the dissolved minerals and the solid residue.
- (f) The correlation model between mineral properties (%Mg, particle size and mineral composition), dissolution parameters (T, P, time, L/S ratio, mineral/additive ratio, etc.) and cost/energy consumption for the dissolution step will be provided. The energy consumption correlation models will be compared and the integration of chemical dissolution stage using NH₄-salts with selected CCSM process will be investigated.

5.2.4 Carbonation reaction (D2.4vii)

The UoN will be investigate the carbonation of UK minerals by assessing:

- a) The reactivity of the UK serpentinite resources in mineral carbonation. Sets of experiments will evaluate the reaction efficiency of selected UK mineral resources. The carbonation rates will be measured as a function of T, additive concentration (or L/S ratio), duration, and CO₂ concentration in gas stream (pure CO₂ and simulated flue gas). The input for a large-scale chemical carbonation reactor design will be provided to CPI/FW.

- b) The correlation model between mineral properties (%Mg, and mineralogical characteristics), carbonation parameters (T, P, time, L/S ratio, etc.) and cost/energy consumption for the carbonation step will be provided. This data will be used for the detailed TEA of the CCSM technology.
- c) The following equipment will be used: TGA-MS, X-Ray Diffraction (XRD), Inductively Coupled Plasma-Mass Spectrometry (ICP-MS), XRF and will be used for the qualitative and quantitative analysis of the carbonation products and un-reacted rocks to establish the effect of the investigated variables on the carbonation kinetics.

The data collected will significantly improve the consortium's understanding of the fundamentals of CCSM and show how to reduce the capital cost of the technology. These experiments will fill the data gap to allow an assessment and initial outline process flows and conditions to be determined for a lower cost more efficient integrated CCSM process. The output will be used for a detailed TEA of the CCSM technology by CPI with Foster Wheeler. The data from these experiments will fill the gaps identified to date and will reduce uncertainty around an integrated CCSM system.

5.2.5 Process Design

The final part of the process requiring further work is the basic process design. This will be undertaken through a combination of routes including the existing Shell Aspen model, but also including outline designs for all processes. It is hoped that improvements in the design of the process plant elements will reduce size and scale of the surrounding equipment and the proposed development activities are targeted at delivering these process improvements. It is likely that this type of approach will reduce process scale and potentially.

6 Conclusions and Specific Technology Development Opportunities

Currently, mineralisation processes would rely on plants that are physically too large to be viable and that are too expensive to justify investment. However, there are indications that very significant improvements in cost and design can be made. The consortium has outlined these detailed proposals in its Stage 2b Development Proposal that has been submitted to the ETI. In summary the proposed next steps are:

- Identify plant elements that are common to all processes and seek to design lower cost components or locate lower cost technology;
- Carry out lab and semi-pilot experiments to improve reagent reactivity and reactor design looking at the three process types identified by the team;
- Work to identify and develop cheaper process design options;
- Redesign the process to reduce parasitic energy consumptions;
- Develop more efficient reactor designs

As discussed earlier much of the equipment required is at a High TRL, however further investigation is required to improve and optimise the processes.

Once the next phase of experimental and design work is completed the consortium will have a much clearer idea of the true capital and operating requirements based on technology that is low

cost but can do the job. It will also have a clearer idea of the process development opportunities that create more compact and lower energy processes.

The lower capital, smaller footprint plant should have more favourable economics and the scale of the improvement can be assessed.

At the end of Stage 2b a further Go/No Go review should be undertaken. If the work is successful and an economically acceptable way forward can be identified work should then start to develop a true continuous pilot scale process that can demonstrate the effective operation of a mineralisation process.

Appendix 1

A1 Mining and Transport CCSM costs

The costs given in the following sections are calculated for the total **annual** CO₂ emission at each industrial site. The costs for quarrying and transport were given in the SG1 report section 7.1 and are used here. The costs applied on the two different scenarios are illustrated in Appendix A1.

BALLANTRAE, SCOTLAND TO FELLSIDE HEAT AND POWER, CUMBRIA

LIZARD, CORNWALL TO ABERTHAW CEMENT WORKS, SOUTH WALES

This example shows the costs for the scenario in SW England. The cost per tonne is actually the same as the previous example as the only difference is the slightly greater distance of the sea transport. However sea transport costs only increase very slowly with distance and only when it is thousands of miles do they become significantly greater.

This example shows the costs of quarrying and transport of rock feed material to the emitter site and return transport of the reacted carbonated material. This produces costs of around £8 per tonne of rock transported and as three tons of rock are required for each tone of CO₂ the cost rises to £24 ton of CO₂ captured.

| | Cost (£/tonne _{CO2}) | Power station (£M) 0.33Mt CO ₂ | Cement plant (£M) 0.56Mt CO ₂ |
|----------------------------------|--------------------------------|---|--|
| Cost of quarrying | 6 | 2 | 3.4 |
| Cost of rock delivery | 9 | 3 (107km) | 5 (181km) |
| Load/unload in delivery | 6 | 2 | 3.4 |
| Cost of transport back to quarry | 12 | 4 (107km) | 6.7 (181km) |
| Load/unload in transport back | 8 | 2.7 | 4.5 |
| Total | 41 | 13.7 | 23 |

A2 GCCS costs

FELLSIDE HEAT AND POWER, CUMBRIA FOR UNDERGROUND STORAGE IN SOUTHERN PART OF NORTH SEA.

ABERTHAW CEMENT WORKS, SOUTH WALES, BRISTOL CHANNEL FOR UNDERGROUND STORAGE IN SOUTHERN PART OF NORTH SEA.

The Fellside Heat and Power produces an emission 0.33Mtpa CO₂ from the 168 MW generator. The Aberthaw Cement Works emits 0.56Mtpa CO₂. For the pipeline costs we use the figures from ZEP 2011 and for monitoring and storage McKinsey 2008. Storage and monitoring cost is for 40 years only.

GCCS pipeline:

| | Cost (£/tonne _{CO2}) | Power station (£M) 0.33Mt CO ₂ | Cement plant (£M) 0.56Mt CO ₂ |
|------------------------|--------------------------------|---|--|
| Onshore pipeline | 6.3 | 2.1 (320km) | 3.53 (354km) |
| Offshore pipeline | 7.4 | 2.44 (100km) | 4,14 (100km) |
| Storage and monitoring | 8 | 2.64 | 4.48 |
| Total | 21.7 | 7.18 | 12.15 |

GCCS tanker:

| | Cost (£/tonne _{CO2}) | Power station (£M) | Cement plant (£M) |
|------------------------|--------------------------------|--------------------|-------------------|
| Liquefaction | 4.21 | 1.4 (1427km) | 2.36 (1164km) |
| Shipping | 11.5 | 3.8 | 6.45 |
| Offshore pipeline | 7.4 | 2.44 (100km) | 4,14 (100km) |
| Storage and monitoring | 8 | 2.64 | 4.48 |
| Total | 31.1 | 10.28 | 1.43 |

Cost sensitivity study (example B), it is assumed that pipeline and storage cost for small scale GCCS is doubled.

| CCS pipeline | Cost (£/tonne _{CO2}) |
|-------------------------------------|--------------------------------|
| Onshore pipeline ⁵ | 12.6 |
| Offshore pipeline | 14.8 |
| Storage and monitoring ⁶ | 16 |
| Total | 43.4 |

| CCS tanker | Cost (£/tonne _{CO2}) |
|------------------------|--------------------------------|
| Liquefaction | 4.21 |
| Shipping | 11.5 |
| Offshore pipeline | 14.8 |
| Storage and monitoring | 16 |
| Total | 46.5 |

A3 Capture Costs

The majority of the costs assessments have been focused on large coal-fuelled power plants for which, the associated avoided cost of CO₂ ranges from \$60 to 150/tonne (£40-90/tonne CO₂) CO₂ (GCCS Institute, 2011; ZEP, 2011; McKinsey, 2010; IMC, 2008).

However, the CCS costs in industrial plants such as refineries, cement and steel works and gas fuelled ammonia plants are less well known and only a few assessments are available, but these indicate relatively higher costs compared to power plants CCS. Also, CCS costs of an early demonstration project (300MW, earliest starting from 2015) will involve higher costs due to their smaller scale, lower efficiency and shorter life (25yrs instead of 40yrs) (McKinsey, 2010). In agreement with the above vision, the report on CCS development and associated costs made by Ian Murray and Co. Ltd (IMC) in 2008 for the Alberta Carbon Capture and Storage Development Council indicates that the costs of capture applied to chemicals, fertilisers, refineries and gas fuelled plants of \$75 to \$235/tonne CO₂ are much higher than costs reported for large centralised emitters. The reasons for the higher costs are related to facility specific issues such as geographic location, production/operating specifics and new technology versus retrofit capture situations (IMC, 2008). Another report focused on industrial emitters and in particular on refineries was the Statoil Hydro master plan for the CCS project at Mongstad. According to initial estimates, the Mongstad project could cost between \$203 and \$280/tonne CO₂. This estimate does not include the transport and storage costs. StatoilHydro reported that the CCS plant is more costly than rival projects because the sequestering will be done on natural gas, which has a lower CO₂ content than coal, and because the plant will need to be integrated with existing facilities at Mongstad (Bellona, 2011).

The capture costs were applied to the two scenarios selected as follow:

| | Fellside heat and power 0.33Mtpa CO₂ | | Aberthaw cement works 0.56Mtpa CO₂ | |
|----------------------|--|-------------|--|-------------|
| | Cost | Actual cost | Cost | Actual cost |
| | (£/tonne _{CO2}) | (£) | (£/tonne _{CO2}) | (£) |
| Capture costs | 121 | 40M | 121 | 68M |

A4 Mineral Carbonation Costs

The available sequestration costs for the CCSM technologies summarised in Section 15 of the first report were applied to the two selected scenarios. Again, it is important to highlight that these assessments were done using different methods and the aim was to establish the wide range of costs involved in the mineralisation technology and not to compare single technologies. The estimated cost of geological sequestration (GCCS) illustrated in 2.2.5 was considered as a base to establish the relative cost of CCSM.

A class D values variation ($\pm 50\%$) was accounted for in the interpretation of the following data, considering all the caveats and the absence of a complete economic evaluation in literature due to the early stage of development of mineralisation technology.

The above value refers to the mineralisation technology but does not include capture stage, post-processing, disposal and transport of material in/out from the conversion site (treated in the previous sections). The capture stage was assumed to be unnecessary considering that both the Shell and CAT processes can work with flue gas. Post-processing was not considered at this point assuming that the CCSM products are transported back at the quarry site for environmental land restoration or disposal.

The mineral carbonation costs were applied to the two scenarios selected as follow:

| | Fellside heat and power 0.33Mtpa CO₂ | | Aberthaw cement works 0.56Mtpa CO₂ | |
|-------------|--|-------------|--|-------------|
| | Cost | Actual cost | Cost | Actual cost |
| | (£/tonne _{CO2}) | (£) | (£/tonne _{CO2}) | (£) |
| CCSM | 300 | 99M | 300 | 168M |

Section 2.4 and 2.5 indicates that GCCS costs are lower when the technology is applied to large emitters while it is more costly when applied to medium small emission sources. Therefore, CCSM can integrate with or replace geological storage in areas far from the North Sea depleted oil and gas fields. It can sequester in permanent way the CO₂ released by small and medium-small industrial plants that present an intrinsic complexity for GCCS (< 300MW) as indicated in the CCSM scenarios. Also, retrofitting a CCS system in an existing power plant will lead to higher costs mainly due to high CAPEX and the installation's shorter life time. CCS costs for industrial plants such as cement and steel works (CCS is around 20% and 42% higher for steel and cement works, respectively) are less well known but indicate higher cost compared to power plants CCS costs (McKinsey, 2010), because the capture of CO₂ is not inherent in the design of

these facilities (GCCSI, 2009). In these specific cases (small industrial emitters) CCSM can play an important role.

A5 GCCS Costs (Geological Storage) Justification

This is a technology still in the demonstration phase, and there are still many questions to be resolved around GCCS before it enters large-scale deployment. One of the most significant of these is the uncertainty around its costs. However, indications are that the range of uncertainty around GCCS is on the same scale as for other emerging technologies. The associated avoided cost of CO₂ ranges from \$60-85/tonne CO₂ for coal based power stations and exceeds \$100/tonne for a gas-fired power plant mainly because natural gas contains lower quantity of CO₂ (<10%) than coal (Global CCS Institute, 2011). In agreement with this vision, the European Technology Platform for Zero Emission Fossil Fuel Power Plants (ZEP) analysis on the costs of deploying the entire GCCS value chain reported for the capture, transport (180km) and storage costs ranging from £37 to £52/tonne CO₂ for 20 and 2.5Mtpa CO₂, respectively (£24-32/tonne for the capture; £4-12/tonne for CO₂ transport; £8/tonne for CO₂ storage)(ZEP, 2011; McKinsey, 2010). Similarly, the GCCSI modelling determined that the cost of CCS for power generation, based on the use of commercially available technology, was found to range from £39 to £72/tonne CO₂ avoided (GCCSI, 2009). However, the above costs were mainly based on large coal power plants but CCS costs in industrial plants have higher costs compared to CCS in large power plants. In agreement with this vision, the report on CCS development and associate costs made by Ian Murray and Co. Ltd (IMC) in 2008 for the Alberta Carbon Capture and Storage Development Council indicates that the costs of CCS applied to chemicals, fertilisers, refineries and gas fuel of \$75 to \$235/tonne CO₂ are much higher than costs reported for large centralised emitters such as coal-fired power plants (\$50 to \$150/tonne CO₂). This is shown in Figure 13 that summarises CO₂ capture costs by major facility type per tonne of CO₂ abated basis (1 Cnd \$ = 1.04 US \$ in 2008, in 2011 Cnd \$ = 1.05, Cnd \$ =£0.64). The range of capture cost for each major facility type is related to facility specific issues such as geographic location, production/operating specifics and new technology versus retrofit capture situations (IMC, 2008).

Another report that focused on industrial emitters and in particular on refineries was the StatoilHydro master plan for the CCS project at Mongstad. According to initial estimates, the Mongstad project could costs between \$203 and 280/tonne CO₂, against a benchmark price of 14.96 euros (\$20.93) per tonne for European carbon emissions purchased on the market. These estimates do not include the transport and storage costs. StatoilHydro said that the CCS plant is more costly than rival projects because the sequestering will be done from natural gas, which has a lower CO₂ content than coal, and because the plant will need to be integrated with existing facilities at Mongstad (Bellona, 2011).

Averagely, the CCS costs can be considered of \$200/tonne CO₂ for industrial emitters.

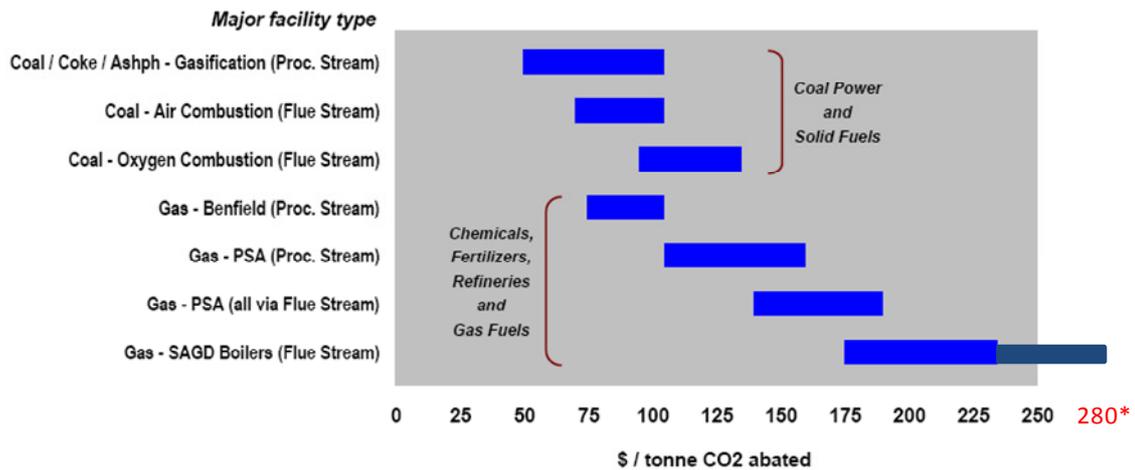


Figure 13. Compilation of Industry CO₂ Capture Cost Estimates, Aggregated Results from Capture Cost Survey (IMC, 2008).

Notes: These figures are based on over 20 different facilities from over 10 company interviews, plus information from other recent studies.- Cost ranges represent geographic, technological suitability as well as greenfield versus retrofit considerations. - \$ are 2008 Cdn, 40 years life, based on leveling real 2008 Capital and real 2008 Annual Operating costs discounted at 10% from year(s) incurred. - \$/tonne abated includes cost penalties for make-up production and incremental CO₂ emissions resulting from CO₂ capture. - Gasification excludes cost penalty, if any, associated with production technology choice relative to alternative(s). - Benfield excludes cost penalty, if any, associated with production technology of choice relative to alternative(s) - IMC standardised some elements of the cost estimates such as the cost of natural gas fuel supplies at \$7 per GJ in the effort to eliminate some of the variances in reported numbers (IMC, 2008). * CCS costs in a refinery with use of natural gas (StatoilHydro, 2009).

The McKinsey report indicates that CCS costs are lower when the technology is applied predominantly to large emitters while it is more costly when applied to medium small emission sources (McKinsey, 2010).

The IMC (2008) costs include all design, construction, owning, operating and maintenance expenditures associated with capture equipment, land and other modifications to the production facility necessary to manufacture dry, high concentration (>95% CO₂), high purity, high pressure (2200 psi), bulk CO₂ ready for injection into a pipeline system. Costs also include all expenditures necessary to make-up any decline in the production facility's primary output that is impacted by capturing CO₂. Typical profit margins for most segments of the CO₂ capture value chain also are recognised as costs, including, for example, margins for engineering suppliers, equipment suppliers, construction service suppliers, operations and maintenance service suppliers, consumables and fuel suppliers. Profit margins, if any, related to owning and operating capture facilities are assumed to be inherent in the 10% pre-tax discount rate used for calculating unitised capture costs (IMC, 2008). Capital and operating costs represent about 70% and 30%, of the total expenditure, respectively (ZEP, 2011; IEA, 2008; Wright et al, 2009).

A6 Transport Costs Justification

The transport costs range from £13 to £41/tonne CO₂ for a distance of 1500km and 20 and 2Mtpa CO₂ transported, respectively (ZEP, 2011). Therefore, the GCCS costs may reach £85/tonne CO₂ if applied to emitters smaller than 2.5Mtpa and far from the storage site. The pipeline costs are roughly proportional to distance with costs varying from £4 to £41/tonne CO₂ for a distance between source and storage site of 180 km and 1500 km, respectively. In contrast,

shipping costs are only marginally influenced by distance (£6 to £13/tonne CO₂). Also, pipeline costs consist mainly (> 90%) of CAPEX, while shipping costs are less CAPEX intense (< 50%). The liquefaction cost required for ship transport is included in the cost but compression for pipelines is not included. The investigation indicates that for CCS to be commercially viable it will have to serve large scale power plant of 20Mtpa leading to a decreased transport cost of about 60% and 20% for pipelines and ships, respectively. Therefore, pipelines benefit from large scale in a major way compared to shipping transport.

A6.1 Transport in pipelines and ships justification

The ZEP report (2010) fits very well to our purpose considering that they studied the transport costs for a demonstration projects with a typical capacity of 2.5Mtpa. Table 6 shows the unit transportation cost for such projects depending on transport mode, distance and volumes transported.

Table 6. Cost Estimates for Demonstration Phase Projects, 2.5Mtpa (Eur/tonne CO₂).

| <i>Distance km</i> | <i>180</i> | <i>500</i> | <i>750</i> | <i>1500</i> |
|-----------------------------------|----------------|--------------|------------|---------------|
| Onshore pipe | 5,4 (£4.29) | n a (£6.36)* | n a | n a |
| Offshore pipe | 9,3 (£7.4) | 20,4 | 28,7 | 51,7 |
| Ship | 8,2 | 9,5 | 10,6 | 14,5 (£11.52) |
| Liquefaction (for ship transport) | 5,3 | 5,3 | 5,3 | 5,3 (£4.21) |

*** assumed from the values reported in the ZEP report for large emitters (20Mtpa) where the transport cost more than double from 180 to 500km (ZEP, 2010).**

As shown, pipeline costs are roughly proportional to distance, while ship costs are only marginally influenced by distance. Considering technical or commercial risk involved in demonstration projects, the construction of a single demonstration project point-to-point pipeline is less attractive than ship transportation at distances below 500km. Pipeline costs here exclude any compression costs at the capture site while the required liquefaction cost for ship transport is specified. At a stage when CCS becomes a commercially driven reality, it is assumed that typical volumes will be in the range of 10Mtpa, serving for instance one large scale power plant or 20Mtpa serving a cluster of sources.

A6.2 Relevant Assumptions

The ZEP cost estimations are likely to be accurate with a margin of about 30%. For short to medium distances and large volumes pipelines are by far the most cost effective solution but the implementation requires strong central coordination.

One of the most important assumptions in the ZEP report is the one about production volumes and profiles. This report considers three different annual volumes, 2.5Mtpa assumed typical for a demonstration project 10Mtpa representing a full scale, commercial coal-fired power plant with CCS and 20Mtpa assumed typical for a full scale mature CCS scheme. Therefore, our assumption was to consider the lower volumes to compare the CCSM technology considering that the target of mineral carbonation is on small emitters far from storage sites. However, it is stressed that the scenarios for CCSM are nearly an order of magnitude smaller, there is no comparable data and hence the actual figures for GCCS for the same size could be substantially higher.

The presence of water is the most critical impurity issue influencing the cost of transport since the risk of corrosion on pipes, tanks and process equipment directly affects the choice of material. For such transport systems, it is not realistic to assume using exotic materials having

resistance to a corrosive CO₂, since this would increase costs by an order of magnitude. The cost estimates in this report are therefore based on the use of normal carbon steel, assuming the carbon dioxide for transport is sufficiently dry. It should be noted that the liquefaction process used for ship cargoes also requires low humidity. These assumption allow us to use the higher values reported due to higher capture costs if no moisture is required.

To avoid numerous compressor stations along the pipeline, the inlet pressure, also for onshore pipelines, probably needs to be higher than the existing maximum requirements for gas pipelines but was not considered in the ZEP costs, so in the real world higher costs are expected.

The currency exchange rates used are

- 0.68 EUR/USD
- 1.26 EUR/GBP
- 0.11 EUR/NOK

The agreed CCS project lifetime is forty (40) years.

After studying European electricity market prices and after an exchange with McKinsey, an electricity price of 0.11 EUR/kWh was found to be representative.

CO₂ carriers (ships) for CCS purposes are likely to be most typically in the 20-30 000 m³ range. In order to reduce the cost of ships and storage tanks (thickness of tanks walls); it is preferable to operate as close to the triple point of -56.6°C/5.2 bar as is practically feasible. Liquefaction of the CO₂ is accomplished by condensing and depressurising. The temperature is controlled by the pressure. Between 20 and 40% of the CO₂ condensate/dense phase will flash off during depressurisation and has to be recompressed.

If a pipeline is installed where there is potential for greater capture in the future, it is likely that planning and environmental impact considerations would be assessed to match the potential future capacity. This is due to the marginal cost difference, say in the case of a 10Mtpa (as in the base case for a capture location) pipeline compared to a 2.5Mtpa pipeline. Therefore pipelines may be sized at greater capacity than the scale of the demonstrator project requires.

Shipping is particularly relevant to the demonstration phase as it has the least chance of being a stranded asset if the project does not continue past the demonstration phase. If a location would never produce a peak capture rate of more than 2.5Mtpa then the associated pipeline would be that size. Therefore, in our costs comparison the shipping option was considered as the most likely option for GCCS at the locations under consideration a long distance from the North Sea.

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