Evaluating technology options to mitigate greenhouse gas emissions

Review of Fuel Cell Technologies with CO_2 Capture for the Power Sector

IEAGHG Technical Report

2019-03

April 2019

IEA GREENHOUSE GAS R&D PROGRAMME

International Energy Agency

The International Energy Agency (IEA), an autonomous agency, was established in November 1974. Its primary mandate was – and is – two-fold: to promote energy security amongst its member countries through collective response to physical disruptions in oil supply, and provide authoritative research and analysis on ways to ensure reliable, affordable and clean energy for its 30 member countries and beyond. Within its mandate, the IEA created Technology Collaboration Programmes (TCPs) to further facilitate international collaboration on energy related topics. To date, there are 38 TCPs who carry out a wide range of activities on energy technology and related issues.

DISCLAIMER

The GHG TCP, also known as the IEAGHG, is organised under the auspices of the International Energy Agency (IEA) but is functionally and legally autonomous. Views, findings and publications of the IEAGHG do not necessarily represent the views or policies of the IEA Secretariat or its individual member countries."

This report was prepared as an account of the work sponsored by IEAGHG. The views and opinions of the authors expressed herein do not necessarily reflect those of the IEAGHG, its members, the organisations listed below, nor any employee or persons acting on behalf of any of them. In addition, none of these make any warranty, express or implied, assumes any liability or responsibility for the accuracy, completeness or usefulness of any information, apparatus, product of process disclosed or represents that its use would not infringe privately owned rights, including any parties intellectual property rights. Reference herein to any commercial product, process, service or trade name, trade mark or manufacturer does not necessarily constitute or imply any endorsement, recommendation or any favouring of such products.

COPYRIGHT

ACKNOWLEDGEMENTS AND CITATIONS

This report describes research commissioned by DOOSAN UK on behalf of IEAGHG. This report was prepared by:

- Jonathan Slater
- Theodore Chronopoulos

To ensure the quality and technical integrity of the research undertaken by IEAGHG each study is managed by an appointed IEAGHG manager. The report is also reviewed by a panel of independent technical experts before its release.

The IEAGHG manager for this report was:

- Monica Garcia Ortega

The expert reviewers, in surnames alphabetical order, for this report were:

- Linda Barelli, University of Perugia
- Moufid BenMerabet, OPEC
- Suzanne Ferguson, Wood UK
- Timothy E. Fout and Gregory Hackett, NETL
- Robin Hughes, CanmetENERGY, Natural Resources Canada
- Angelo Moreno, Atena- Future Technology
- Robert Steinberger-Wilckens, University of Birmingham

The report should be cited in literature as follows:

'IEAGHG, "Review of Fuel Cell Technologies with CO_2 Capture for the Power Sector", 2019/03, April, 2019.'

Further information or copies of the report can be obtained by contacting IEAGHG at:

IEAGHG, Pure Offices, Cheltenham Office Park,
Hatherley Lane, Cheltenham, Glos. GL51 6SH, UK
Tel: +44 (0)1242 802911
E-mail: mail@ieaghg.org Web: www.ieaghg.org

REVIEW OF FUEL CELL TECHNOLOGIES WITH CO_2 CAPTURE FOR THE POWER SECTOR

Key Messages

- This study aims to provide a techno-economic overview of the most promising fuel cells (FCs) for the power sector. It also identifies future research areas to enhance their performance and describes barriers and challenges to achieve commercial deployment
- The fuel cells assessed are SOFCs (Solid Oxide Fuel Cells) and MCFCs (Molten Carbonate Fuel Cells) with and without further CO_2 capture/purification, in Supercritical Pulverised Coal (PC), Natural Gas Combined Cycle (NGCC, also known as CCGT, Combined Cycle Gas Turbine) and Integrated Gasification Combined Cycle (IGCC) power plants
- The identified typical CO_2 capture/purification systems to use in combination with FCs are: cryogenic CO_2 separation, oxy-combustion, oxidation[1], condensation[2], membrane separation and physical absorption
- Two generations of SOFCs are identified, with the second generation offering promising cost reductions, although they are currently at low TRL (Technology Readiness Level) and therefore require an initial higher investment. For MCFCs, the general practice is to use those as an “active CO_2 concentrator” in combined cycles
- To facilitate the comparison of the costs reported in the literature, those are homogenised by currency, year, plant size and location. The results are compared with reference costs from cases without and with a benchmark CO_2 capture system (chemical absorption with Cansolv solvent)
- After homogenisation, SOFCs and MCFCs are shown as competitive systems for CO_2 capture in power plants
- The lowest CO_2 avoidance cost is showed by the MCFC case operating at atmospheric pressure in a NGCC (Natural Gas Combined Cycle) plant, with a CO_2 capture system using oxy-combustion followed by condensation.
- The lowest LCOE (levelized cost of electricity[3]) is shown by the MCFC case operating at atmospheric pressure in an integrated NGCC with CO_2 purification with cryogenic separation. This case represents a reduction of the LCOE by 24% compared to the use of the benchmark CO_2 capture system.
- The dependency of the LCOE on the fuel cells key parameters was assessed through a sensitivity analysis in the case of a NGCC with a MCFC operating at atmospheric pressure. The fuel cell investment was found to have the highest impact on the LCOE. Degradation[4] is identified in the literature as key parameter and it is studied as a variation on the stack life. However, in this study, its effect was less significant

---

[1] During oxidation, the remaining fuel is oxidised in high-purity O_2

[2] During condensation, the water content is separated by temperature difference

[3] LCOE is an appropriate comparative tool given all power plants compared here are baseload. To note that this metric does not take into account the value of different technology features and system dependencies.

[4] Degradation is defined here as certain power losses associated to carbon deposition due to the lack of a water-gas-shift (WGS) step

- Challenges and barriers for the commercial deployment of fuel cells technologies with CO_2 capture can be divided into manufacturing and materials, operation, public policies and the need for large-scale projects.

Background to the Study

CO_2 emissions from the energy production sector were estimated as 42% of the global share in 2016. In order to achieve the below 2°C scenario, the emissions from the power sector should be reduced from 40 to 4.7 $GtCO_2$ by 2050[5] (IEA, 2017).

Carbon capture systems are needed within the mix of measures for decarbonisation. Pre-, oxy- and post-combustion technologies appear as advanced systems, but the energy penalty and consequent efficiency drop are still large concerns.

Electrochemical conversion with fuel cells has been proposed as a more efficient carbon capture option. Two types of fuel cells, Molten Carbonate Fuel Cells (MCFCs) and Solid Oxide Fuel Cells (SOFCs) have recently emerged as alternatives to capture CO_2 with the added advantage of additional power production.

Although some information on electrochemical conversion was included in the review of emerging CO_2 capture technologies delivered by IEAGHG (2014), it was identified the need of a further techno-economic analysis on the most promising fuel cells for the power sector, MCFCs and SOFCs.

Scope of Work

DOOSAN U.K. was commissioned by IEAGHG to provide a comprehensive techno-economic review on MCFCs and SOFCs for the power sector. The objectives of this technical study were:

- Deliver a literature review on MCFCs and SOFCs, identifying available configurations, status of development, applications and gaps to reach economically viable solutions
- Supported by data from the literature, provide a techno-economic evaluation on selected cases and compare to a number of reference cases with and without a benchmark CO_2 capture system (chemical absorption with Cansolv technology)
- Identify key parameters and areas impacting the price of implementing those technologies in the power sector as CO_2 abatement systems
- Describe barriers and challenges to be addressed for SOFCs and MCFCs to achieve commercial application

---

[5] Comparison of RTS (Reference Technology Scenario) and B2DS (Beyond 2°C Scenario) provided in IEA(2017)

Findings of the Study

- Literature review

Fuel Cells (FCs) convert the chemical energy of a gaseous fuel into electricity and heat. A schematic figure with several fuel cells options is included below (US DoE, 2004).

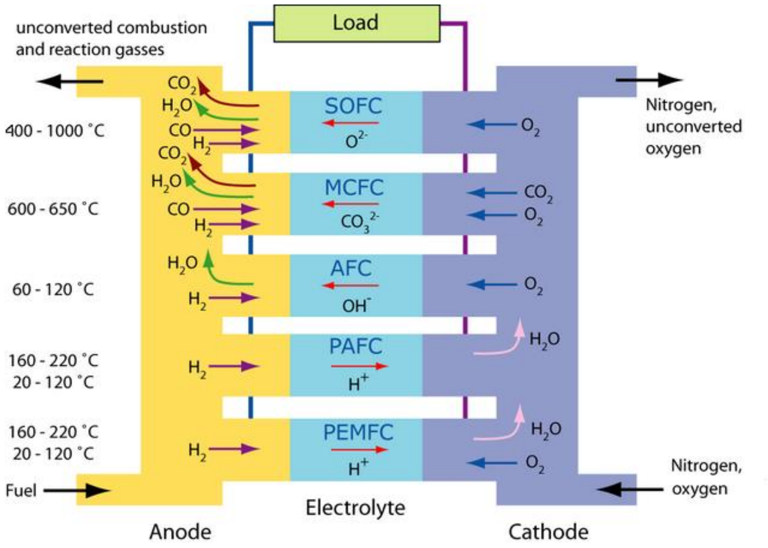


Figure 1: Configuration of several FCs and their operating temperature (US DoE, 2004) (SOFC: Solid Oxide Fuel Cell; MCFC: Molten Carbonate Fuel Cell ; AFC: Alkaline Fuel Cell; PAFC: Phosphoric Acid Fuel Cell; PEMFC: Proton Exchange Membrane Fuel Cell)

CO_2 capture can take place at different process stages such as after the FC (post-anode) or upstream of the FC (pre-anode). SOFCs and MCFCs are particularly interesting for power plants applications due to their high operation temperature, which allow the reforming and shift reactions in the fuel cell itself. Based on their high electrical efficiencies and the ability for cogeneration of electricity and heat, MCFCs and SOFCs are suitable for stationary power production.

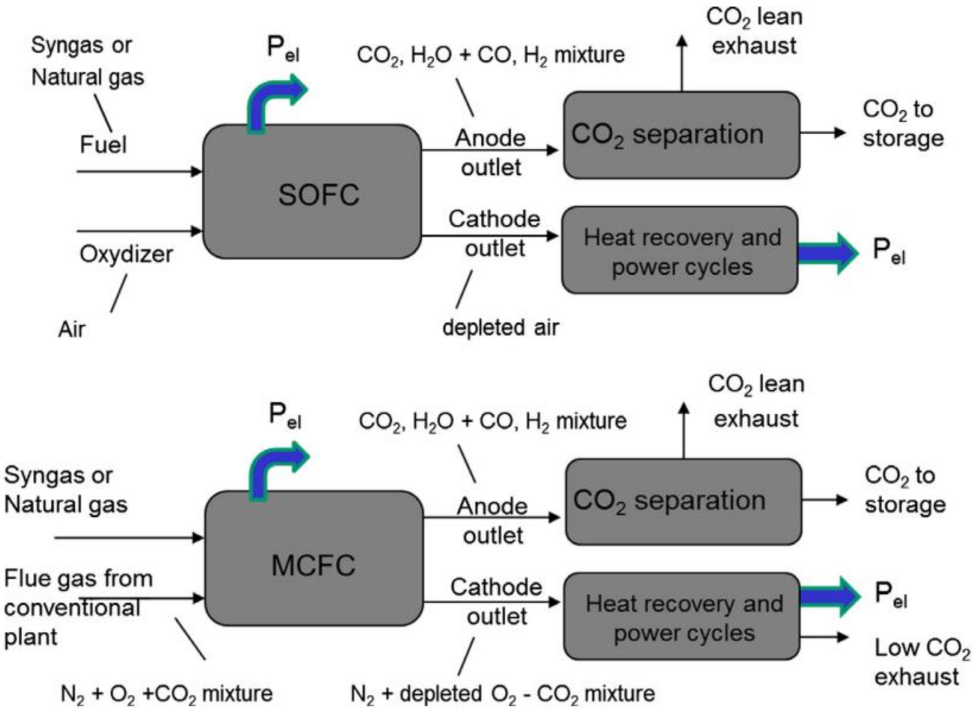


Figure 2 Two typical configurations using SOFC (top figure) and MCFC (bottom figure) (IEA, 2008)

SOFCs allow the conversion of different fuels and operate at relatively high temperature (700-1000°C), although recent advances show operations at lower temperature, covering the 400-1000°C range. In the fuel cell, the water gas shift and steam reforming reactions take place. Alternatively, it is possible to place a water gas shift reactor prior the fuel cell. The hydrogen is oxidized in the cathode while the oxygen is reduced in the anode.

The first generation of SOFCs, operating at atmospheric pressure, is showing promise for commercialization in a near future, based on its advance development status and the tests carried out at large scale. The second generation, at pressurized conditions, has the potential to enhance the overall power plant performance. However, the added complexity and cost must be studied in an individual basis to assess the net benefits.

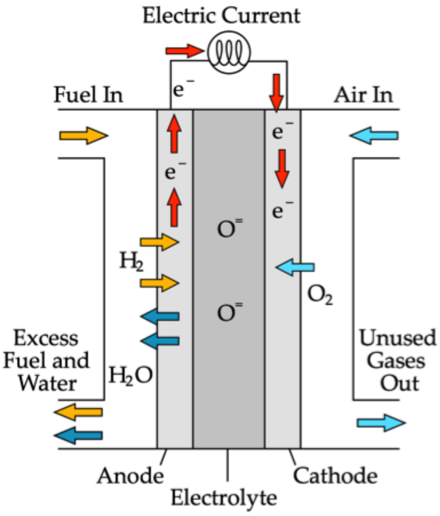


Figure 3: SOFC description (US DoE, 2004)

MCFCs consist on individual cells staked to achieve any desired power, operating typically at 550-650°C. MCFCs can operate with hydrocarbons (shift reaction) or pure H_2, and utilizes carbonate ions as charge carriers for the cathode side. CO_2 is needed for the operation of the MCFCs as a closed-loop reagent (the consumption at the cathode occurs at the same rate of production at the anode). The design parameters are the stack size, heat transfer rate, voltage, loading, and cost. The performance will be determined by the cell pressure, temperature, gas composition and fuel utilization (Uf), based on the Boudouard reaction, methanation and CH_4 decomposition reactions. Additionally, impurities will impact the fuel cell operation.

MCFCs can simultaneously act as a 'CO_2 concentrator' and a 'power producer'. The concentration of CO_2 in the anode outlet is around 80% (dry basis) but this is also dependent on the fuel cell voltage. The CO_2 concentration will be higher in operations at higher voltage, but techno-economically limited due to degradation over the time. Further CO_2 concentration can be obtained, for example, through cryogenic separation, oxy-combustion.

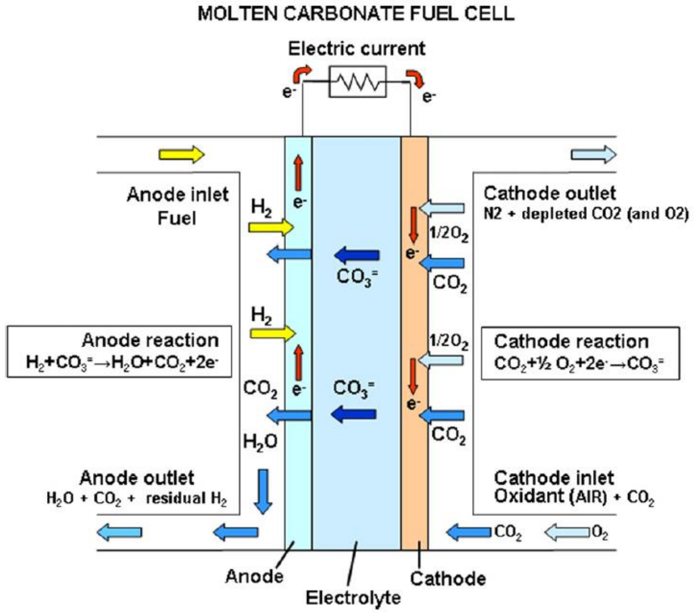


Figure 4 MOFC description (Caprile, et al., 2011)

- **Study basis**

A screening process was carried out to select transparent cost assessments available in the literature. Selected cases were scaled-up to a net power of 634MWe following a linear regression on a €/KW_e basis. Costs were updated on currency (2017 Euros) and plant location (Netherlands, based on the location factors from IEAGHG (2018)). While some of the economic parameters are common to the IEAGHG studies, other assumptions are based on the literature and contractor experience.

For each selected case, installed costs, LCOE and cost of CO_2 avoided were calculated, excluding transport and storage costs and without including any CO_2 tax or CO_2 captured value. Due to the impact of the economic parameters and assumptions, a sensitivity analysis was performed to assess the influence of the fuel cell investment cost, stack life and stack replacement cost on the LCOE. Operating parameters such as: maintenance, chemicals and consumables and labour costs, were evaluated. Moreover, economic parameters which could be region-specific were also assessed, such as the fuel price, discount rate, insurance and taxes, contingency, and interest during construction.

- **Techno-economic assessment of cases from the literature**

The technical and economic assessments were based on data from the literature. The first step was a screening process to select the most transparent cases from the available literature and extract the capital costs. The second step was to establish a common bottom-up cost method to assess the selected cases for a Western European location, for a unique plant size and updated currency and year. The capital costs used are shown as "Total Plant Cost" (TPC), which comprises the Bare Erected Cost (BEC, divided as 60% Equipment cost, 30% Labour cost and 10% Material costs) plus the cost of services provided by the engineering, procurement and construction (EPC) contractor (such as detailed design, permitting and project management, defined as 7.5% of BEC) and project and process contingencies (10% of BEC).

Eleven cases were selected from the literature, six representing reference plants and five assessing SOFCs and MCFCs in power plants, as described in the table below.

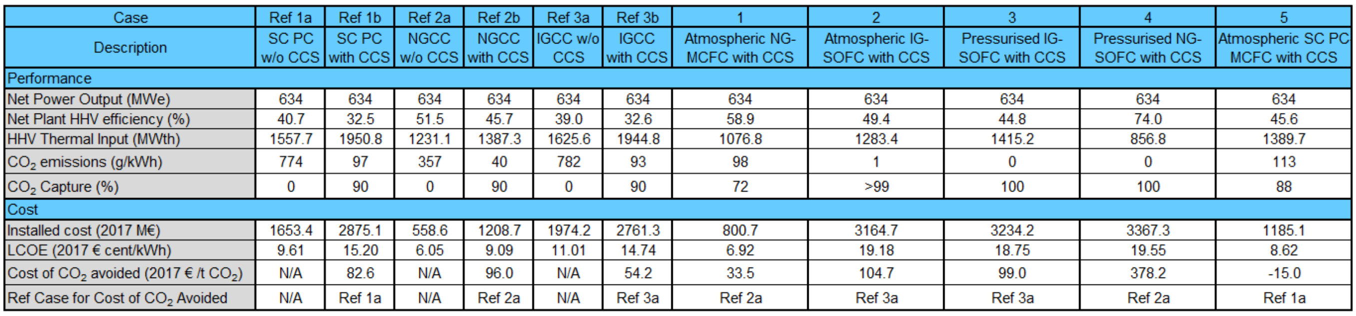
Table 1 FCs cases selected from the literature and included in the homogenised techno-economic review

Case	Description
Reference Case 1a	Supercritical Pulverised Coal (PC) boiler plant without CCS (NETL, 2015)
Reference Case 1b	Supercritical Pulverised Coal (PC) boiler plant with CCS (Cansolv CO_2 Capture Process) (NETL, 2015)
Reference Case 2a	Natural Gas Combined Cycle (NGCC) plant without CCS (NETL, 2015)
Reference Case 2b	Natural Gas Combined Cycle (NGCC) plant with CCS (Cansolv CO_2 Capture Process) (NETL, 2015)
Reference Case 3a	Integrated Gasification Combined Cycle (IGCC) plant (GEE Radiant Gasifier) without CCS (NETL, 2015)
Reference Case 3b	Integrated Gasification Combined Cycle (IGCC) plant (GEE Radiant Gasifier) with CCS (2-Stage Selexol CO_2 Capture Process) (NETL, 2015)
1	Atmospheric integrated NGFC (utilising MCFC) with CO_2 purification through cryogenic CO2 separation. (Campanari, et al., 2014)
2	Atmospheric IGFC (utilising SOFC) with CO_2 capture through oxy-combustion followed by condensation. (NETL, 2009)
3	Pressurised IGFC (utilising SOFC) with CO_2 capture through oxidation followed by flash cascade. (Adams & Barton, 2010)
4	Pressurised integrated NGFC (utilising SOFC) with CO_2 purification through oxidation followed by flash cascade. (Adams & Barton, 2010)
5	Atmospheric MCFC with Natural Gas at the anode and exhaust from PC boiler at the cathode, with CO_2 capture through oxy-combustion followed by condensation. (Duan, et al., 2016)

To note that the reference power plant efficiencies of the PC, NGCC and IGCC power plants (cases 1a,2a and 3a) are 40.7, 51.5 and 39% (HHV basis) respectively (see Table 2). The reference cases with CO_2 capture systems (1b,2b and 3b) consider a capture rate of 90% and their efficiencies are 32.5, 45.7 and 32.6 % (HHV basis) respectively. Moreover, the CO_2 avoidance cost was calculated following the formula below:

$$Cost\ of\ CO_2\ Avoided = \frac{(\,LCOE_{\text{CCS}} - LCOE_{\text{Non-CCS}})}{(\,CO_2\ Emissions_{\text{Non-CCS}} - CO_2\ Emissions_{\text{CCS}})}$$

Table 2 Summary of key techno-economic results from this study



Case	Ref 1a	Ref 1b	Ref 2a	Ref 2b	Ref 3a	Ref 3b	1	2	3	4	5
Description	SC PC w/o CCS	SC PC with CCS	NGCC w/o CCS	NGCC with CCS	IGCC w/o CCS	IGCC with CCS	Atmospheric NG-MCFC with CCS	Atmospheric IG-SOFC with CCS	Pressurised IG-SOFC with CCS	Pressurised NG-SOFC with CCS	Atmospheric SC PC MCFC with CCS
Performance											
Net Power Output (MWe)	634	634	634	634	634	634	634	634	634	634	634
Net Plant HHV efficiency (%)	40.7	32.5	51.5	45.7	39.0	32.6	58.9	49.4	44.8	74.0	45.6
HHV Thermal Input (MWth)	1557.7	1950.8	1231.1	1387.3	1625.6	1944.8	1076.8	1283.4	1415.2	856.8	1389.7
CO_2 emissions (g/kWh)	774	97	357	40	782	93	98	1	0	0	113
CO_2 Capture (%)	0	90	0	90	0	90	72	>99	100	100	88
Cost											
Installed cost (2017 M€)	1653.4	2875.1	558.6	1208.7	1974.2	2761.3	800.7	3164.7	3234.2	3367.3	1185.1
LCOE (2017 € cent/kWh)	9.61	15.20	6.05	9.09	11.01	14.74	6.92	19.18	18.75	19.55	8.62
Cost of CO_2 avoided (2017 € /t CO_2)	N/A	82.6	N/A	96.0	N/A	54.2	33.5	104.7	99.0	378.2	-15.0
Ref Case for Cost of CO_2 Avoided	N/A	Ref 1a	N/A	Ref 2a	N/A	Ref 3a	Ref 2a	Ref 3a	Ref 3a	Ref 2a	Ref 1a

As seen in the Table 2, the MCFCs configurations at atmospheric conditions, in an integrated NGCC with MCFC and further CO_2 purification through cryogenic separation and using natural gas at the anode and PC boiler exhaust gas at the cathode with CO_2 capture through oxy-combustion and condensation after (cases 1 and 5), are the ones showing the lowest LCOE and CO_2 avoidance cost respectively, lower than the reference plants with Cansolv-based post-combustion capture but with lower capture rates than the references. Those lower costs are based on lower installed costs and higher net plant HHV efficiency.

The case 4, pressurised integrated NGCC with SOFC and CO_2 purification by oxidation and flash cascade, appears as the most expensive configuration due to a high installation cost (4 and 3 times the installed costs of cases 1 and 5 respectively).

- **Challenges**

Although fuel cells have been shown to be competitive, there are still economic challenges. For the most part, CCS has not progressed beyond the demonstration stage, with stakeholders hesitant to commit the resources necessary to scale-up and roll-out deployment of the technology. This hesitant progress reflects several characteristics: the large financial magnitude of individual investments, the creation of new networks, the relatively high level of perceived technology and real commercial and policy risks, together with the often tentative commitments of governments to deep decarbonisation investments. The implementation of large-scale plants with fuel cells for CO_2 capture would reduce the technical and financial risks, and increase the confidence of future investors.

In addition to several existing financing mechanisms, understanding the value proposition of large scale deployment of CCS is key. In particular, governments may wish to determine the future role of CCS within their own decarbonisation trajectories and explore how investments now might avoid larger decarbonisation costs later (associated with deployment of less cost-effective technologies). The role of CCS in avoiding large decarbonisation costs in the long term is a key aspect of the value of CCS, particularly from the point of view of consumers. CCS enables access to large amounts of low carbon energy, reducing prices of a large amount of consumer goods compared to a decarbonisation scenario with no CCS.

Expert Review Comments

Comments on the draft were received from seven reviewers in the fuel cells industry and research and energy policy organisations.

A general view of the reviews was that the report was understandable and provided a good contribution to the knowledge in the subject area. Key suggestions included a request for more information on the economic analysis and decisions on the performance of the different cases assessed, along with various other detailed comments. The contractor addressed most of the comments and included further clarification of some aspects. Comments which were not addressed and the reasons to do so are summarized as follows:

- Several reviewers questioned the performance and price of fuel cells considered in the cases assessed. As observed in Table 2, different capture rates are considered. As observed in the Annexes to the main report, the FCs prices change from one case to another. As this study is a review of the available literature and the aim is to homogenise the costs under the same economic framework, the performance and original prices reported in the literature were kept. Capital costs were extracted from the selected studies and homogenized by year, location and size.

- It must be mentioned that the studies were selected based on a screening process, which evaluated the transparency of the information and the technologies included. However, the contractors and IEAGHG recognise that the accuracy of the economic calculations is a limitation of the study due to the use of different sources of information. This study was not intended to include modelling work, because it is an attempt to assess the cases with seemingly the same amount of rigour. It must be understood that in the absence of commercial projects, obtaining more accurate figures is a challenge. In an attempt to provide some insight on the dependency of costs on the key parameters, a sensitivity analysis was performed and is included in the main report.

- One reviewer wondered about the stack replacement. The contractor assumed a stack replacement cost at 25% of the fuel cell cost given in the techno-economic analysis. The stack is replaced every 5 years[6], meaning 5 times in the course of 30 years, which equates to 125% of the fuel cell cost.

- Two reviewers asked separately for the CO_2 storage and utilization cost but those are out of the scope of this study

---

[6] Dependent on the fuel cell degradation rate

Conclusions

The results from this study show that FCs have the economic and technical potential to compete with current benchmark technologies for CO_2 capture. Although due to the lack of demonstration projects which can give realistic inputs to the cost estimations those results have a significant uncertainty, this study homogenized costs figures in the literature for different fuel cells configurations.

MCFC configurations at atmospheric conditions, in an integrated NGCC with further CO_2 purification through cryogenic separation (case 1) and using natural gas at the another and PC boiler exhaust gas at the cathode with CO_2 capture through oxy-combustion and condensation after (case 5) are the options at lowest cost.

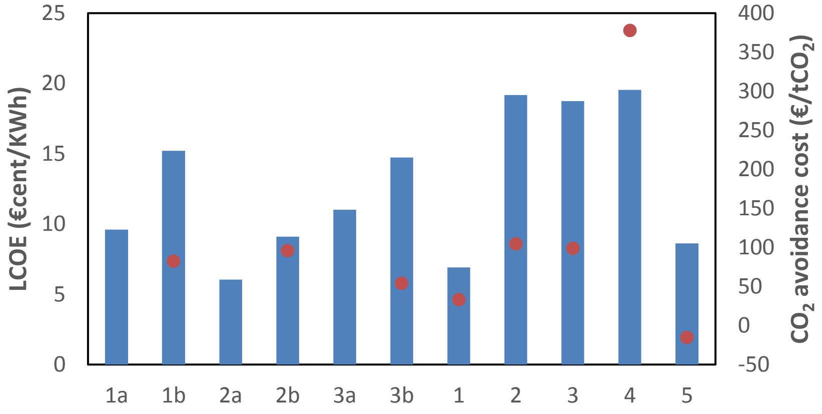


Figure 5 Levelized cost of electricity (LCOE) (blue bars) and CO_2 avoidance cost (red dots) of MCFCs and SOFCs cases described in Table 1

In this study, the fuel cell investment cost has the highest impact on the LCOE and CO_2 capture cost. The cost of the FCs can be divided into material and component costs, labour, and the capital cost of manufacturing. Labour, capital and manufacturing costs can be reduced through economies of scale in manufacturing. However, material and component costs are dependent on technological innovations and the market. For the other components, steel, nickel, zirconia and other materials could impact the stack cost but that can also decrease through manufacturing cost reductions. Utilities and other components contribute to a large portion.

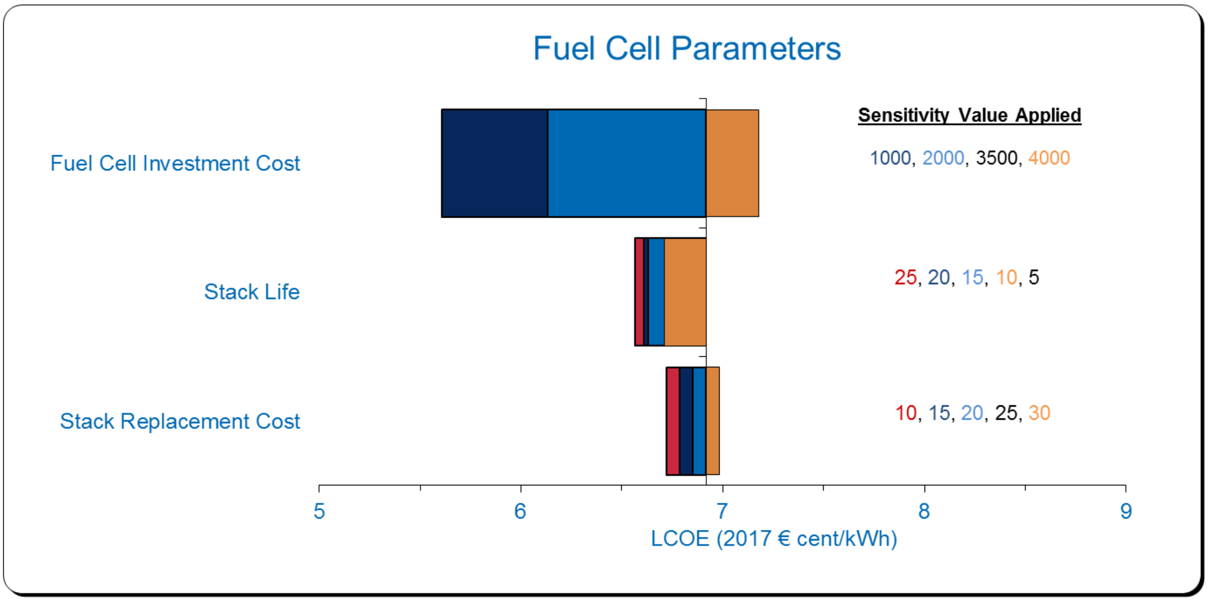


Figure 6: Case 1 Atmospheric NGFC-MCFC with CO_2 Capture: LCOE Sensitivity to FC Parameters

The fuel cell stack life was identified in the literature as a significant parameter impacting on LCOE and CO_2 capture cost. However, in this study, it was observed that this contribution is less significant than the fuel cell investment cost.

Working under pressurized or atmospheric conditions will impact on the costs and must be considered as a design parameter. While an increase on the working pressure will increase the efficiency, that will also add some complexity and additional installed costs due to added auxiliary equipment and more expensive materials.

In terms of operational challenges, pressure management is key for a safe operation, where materials and pressure control will be key. Moreover, if further purification is done through oxy-combustion, oxygen handling will demand some attention. Costs will be influenced by the FC performance, which is affected by operation parameters such as the current density, gas composition, fuel utilisation, pressure and temperature.

Although this study shows that FCs are promising solutions to tackle CO_2 emissions in the power sector, the high investment, operation costs and auxiliaries contribute to a high LCOE and CO_2 capture cost. Challenges and barriers can be split into: manufacturing and materials, operation, public policy and large scale CCS deployment. Based on the early stage of commercialization of FCs, those systems still require financial support (e.g. grants, enhanced depreciation allowances, feed-in tariffs and fuel discounts). Policy mechanisms are essential and robust and comprehensive regulatory frameworks will promote the CCS deployment.

Recommendations

It is recommended that IEAGHG should continue to maintain a watching brief of fuel cells with CO_2 capture/separation technologies for the power sector as part of its portfolio of emerging CCS technologies.

A further techno-economic assessment should be undertaken, including modelling from scratch and where, ideally, operational experience from large-demonstration projects could be incorporated if available It is recommended to combine the results from the present study with the results from "Effects of Plant Location on the Costs of CO_2 capture" (IEAGHG 2018/04) to provide an overview of regions where fuel cells technologies could be key to achieve significant decarbonisation of the fossil fuel power sector.

References

Adams, T. A. & Barton, P. I., 2010. High-efficiency power production from coal with carbon capture. AIChE J., Volume 56.

Campanari, S., Chiesa, P., Manzolini, G. & Bedogni, S., 2014. Economic analysis of CO2 capture from natural gas combined cycles using Molten Carbonate Fuel Cells. Applied energy, Volume 130, pp. 562-573.

Caprile, L., Passalacqua, B. & Torazza, A., 2011. Carbon capture: Energy wasting technologies or the MCFCs challenge?. International Journal of Hydrogen Energy, Issue 36, pp. 10269-10277.

Duan, L. et al., 2016. Study on coal-fired power plant with CO2 capture by integrating molten carbonate fuel cell system. Energy, 117(2), pp. 578-589.

IEA (2008), CO_2 Capture and Storage: A Key Carbon Abatement Option, OECD Publishing, Paris, https://doi.org/10.1787/9789264041417-en. IEAGHG, 2014. Assessment on emerging technologies and their potential to reduce costs, 2014-TR4, December 2014

IEA (2017), Energy Technology Perspectives, interactive graph visited in April 2019, https://www.iea.org/etp/explore/

IEAGHG, 2018. Effects of Plant Location on the Costs of CO2 Capture, 2018-04, April 2018

NETL, 2015. Cost and Performance Baseline for Fossil Energy Plants Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity Revision 3, s.l.: DOE/NETL.

NETL, 2015. Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b - Year Dollar Update, s.l.: DOE/NETL.

US DoE, 2004. Fuel Cell Handbook (Seventh Edition) by EG&G Technical Services, Inc., November 2004 https://www.osti.gov/

Corporate Research and Development Institute (UK)

Title: Review of Fuel Cell Technologies with CO_2 Capture for the power sector

Author(s): Jonathan Slater & Theodore Chronopoulos

Report No :	RD-18-008	Customer :	IEAGHG
Issue:	1	Order No.:	IEA/CON/17/248
Project/Sub-Proj :	108099	Report Type :	R&D
Date of Issue :	18th April 2019		

AMENDMENT CONTROL

Issue	Section	Reason for Amendment
1	All	First Issue

Author :	J. D. Slater	Date:	25th February 2019
Reviewed by :	R. S. Panesar	Date:	25th February 2019
Reviewed by :	F. D. Fitzgerald	Date:	25th February 2019
Approved for Issue by	S. C. Mitchell	Date:	25th February 2019

SUMMARY

Background

At COP 21 in December 2015, 195 countries adopted the first-ever universal, legally binding global climate deal (The Paris Agreement). The agreement sets out a global action plan to put the world on track to avoid dangerous climate change by limiting global warming to “well below 2°C” above pre-industrial levels.

In a “well below 2°C scenario”, it is projected that net zero carbon emissions would be required early in the second half of this century. In that case Carbon Capture and Storage (CCS) is considered an integral part of the decarbonisation strategy mix, since it is able to reduce CO_2 emissions from fossil fuel fired power plants by at least an order of magnitude.

Currently, the most advanced CO_2 capture technologies are:

- Post-combustion
- Pre-combustion
- Oxyfuel combustion

However, the main drawback of the above technologies is the fact that they require large amounts of the energy in order to capture the CO_2. This results in the so called “energy penalty”, resulting in high Levelised Cost of Electricity (LCOE) and power plant efficiency losses.

Innovative capture technologies using high temperature fuel cells that may allow for power to be generated while capturing CO_2, could potentially offer competitive advantages over the above CO_2 capture technologies. The concept of using fuel cells (FCs) within large scale power cycles for CO_2 capture has been studied in greater detail over the last few years, typically focusing on Solid Oxide Fuel Cells (SOFCs) and Molten Carbonate Fuel Cells (MCFCs).

SOFCs and MCFCs present possible commercial solutions for the arrangement of low CO_2 emissions fossil fuel plant configurations. The aim of this study is to investigate fully the technological and economic status of these innovative CO_2 capture concepts. Improvements to current CO_2 capture rates, overall plant efficiency at a minimum cost and potential development routes to commercialisation will be investigated while any challenges and barriers will be presented.

Study Description / Approach

A review of fuel cell technology with CO_2 capture was undertaken. The aim of this review was to identify the technical and economic viability of utilising Fuel Cells (FC) as a next generation CO_2 capture technology. As such, the review focused on the feasibility of using those systems in the short- to mid-term future. Both Solid Oxide Fuel Cells (SOFCs) and Molten Carbonate Fuel Cells (MCFCs) were considered as potential FC technologies.

The literature review identified a number of large scale applications integrating SOFCs and MCFCs in various power cycles configurations with CO_2 capture. Two generations of SOFC technologies were identified, namely, first generation SOFC systems that are the simplest option to commercialisation at large scale in the short term, and second generation SOFC systems at high pressure and with separate anode and cathode exhausts. It was concluded that even though 2nd generation systems require higher capital investment, these do have the potential to capture and compress CO_2 at reduced costs in the future, however, this technology is still at low Technology Readiness Levels (TRLs).

On the other hand, one of the most promising configurations reviewed was the use of MCFCs as ‘active CO_2 concentrator’ in combined cycles. The MCFC is placed downstream of the power generation cycle and before the heat recovery steam generator (HRSG), to concentrate the CO_2 from the exhaust while generating electricity. FuelCell Energy has demonstrated this technology in their Combined Electric Power and Carbon-dioxide Separation (CEPACS) project. The most significant findings of this project were reviewed and are presented in this report.

Both SOFCs and MCFCs provide an alternative solution for CO_2 capture applications, however, there are still significant barriers prior to commercialisation particularly with regards to cost.

Results and Discussion

Cases from the literature review were used in a techno-economic analysis of fuel cell arrangements with CO_2 capture options. Potential configurations obtained in the literature review were updated and assessed using a homogeneous approach to allow for a fair comparison of the techno-economic performance of SOFCs and MCFCs as CO_2 capture systems in power plants against a number of benchmark cases.

This study does not include any performance modelling, so the techno-economic modelling was a review and update of studies carried out previously. The techno-economic cases from the literature review were selected on the basis of the studies that gave the most transparent data to enable an update to be carried out with underlying assumptions understood.

Within the cases selected from literature, an attempt was made to cover the most promising configurations of both SOFCs and MCFCs whilst also allowing a comparison of different capture methods and operating conditions (e.g. atmospheric pressure vs. pressurised operation) to be made.

As no modelling was carried out the performance data presented in the studies selected for analysis were taken to apply, regardless of unit sizing. The baseline costs presented were also taken and then are rationalised and homogenised to ensure that a fair comparison could be made between different configurations and the selected FC hybrid cycle arrangements.

The cases considered were as follows:

Case	Description
Reference Case 1a	Supercritical Pulverised Coal (PC) boiler plant without CCS (NETL, 2015)
Reference Case 1b	Supercritical Pulverised Coal (PC) boiler plant with CCS (Cansolv CO_2 Capture Process) (NETL, 2015)
Reference Case 2a	Natural Gas Combined Cycle (NGCC) plant without CCS (NETL, 2015)
Reference Case 2b	Natural Gas Combined Cycle (NGCC) plant with CCS (Cansolv CO_2 Capture Process) (NETL, 2015)
Reference Case 3a	Integrated Gasification Combined Cycle (IGCC) plant (GEE Radiant Gasifier) without CCS (NETL, 2015)
Reference Case 3b	Integrated Gasification Combined Cycle (IGCC) plant (GEE Radiant Gasifier) with CCS (2-Stage Selexol CO_2 Capture Process) (NETL, 2015)
1	Atmospheric integrated NGFC (utilising MCFC) with CO_2 purification through cryogenic CO_2 separation. (Campanari, et al., 2014)
2	Atmospheric IGFC (utilising SOFC) with CO_2 capture through oxy-combustion followed by condensation. (NETL, 2009)
3	Pressurised IGFC (utilising SOFC) with CO_2 capture through oxidation followed by flash cascade. (Adams & Barton, 2010)
4	Pressurised integrated NGFC (utilising SOFC) with CO_2 purification through oxidation followed by flash cascade. (Adams & Barton, 2010)
5	Atmospheric MCFC with Natural Gas at the anode and exhaust from PC boiler at the cathode, with CO_2 capture through oxy-combustion followed by condensation. (Duan, et al., 2016)

To rationalise the data, scaling factors for contingency, material, labour (productivity and cost), equipment cost, currency conversion and currency inflation were used. These rationalised costs were then scaled linearly on a €/kW$_e$ basis so that each case gave a plant with a net power output of 634 MW$_e$.

For each case considered, the levelised cost of electricity (LCOE) was calculated by modelling the net present value (NPV) of the plant's cash flow and adjusting the assumed electricity price stated in the model until it gives a zero NPV. The data for each case was rationalised using the same method and cash flows modelled based on the same economic parameters. The primary value of this approach lies not in the absolute accuracy of cost estimates for the individual cases, but in the fact that all cases were evaluated using a common methodology with a consistent set of economic assumptions to ensure a fair comparison can be made between the various technologies. The calculated LCOE is presented in Euro cent/kWh (€ cent/kWh). The results of the techno-economic analysis were as included in Table 1.

Table 1: Summary of Key Techno-Economic Analysis Results: Reference Cases and Cases 1-5

Case	Ref 1a	Ref 1b	Ref 2a	Ref 2b	Ref 3a	Ref 3b	1	2	3	4	5
Description	SC PC w/o CCS	SC PC with CCS	NGCC w/o CCS	NGCC with CCS	IGCC w/o CCS	IGCC with CCS	Atmospheric NG-MCFC with CCS	Atmospheric IG-SOFC with CCS	Pressurised IG-SOFC with CCS	Pressurised NG-SOFC with CCS	Atmospheric SC PC-MCFC with CCS
Performance											
Net Power Output (MWe)	634	634	634	634	634	634	634	634	634	634	634
Net Plant HHV efficiency (%)	40.7	32.5	51.5	45.7	39.0	32.6	58.9	49.4	44.8	74.0	45.6
HHV Thermal Input (MWth)	1557.7	1950.8	1231.1	1387.3	1625.6	1944.8	1076.8	1283.4	1415.2	856.8	1389.7
CO_2 emissions (g/kWh)	774	97	357	40	782	93	98	1	0	0	113
CO_2 Capture (%)	0	90	0	90	0	90	72	>99	100	100	88
Cost											
Installed cost (2017 M€)	1653.4	2875.1	558.6	1208.7	1974.2	2761.3	800.7	3164.7	3234.2	3367.3	1185.1
LCOE (2017 € cent/kWh)	9.61	15.20	6.05	9.09	11.01	14.74	6.92	19.18	18.75	19.55	8.62
Cost of CO_2 avoided (2017 € /t CO_2)	N/A	82.6	N/A	96.0	N/A	54.2	33.5	104.7	99.0	378.2	-15.0
Ref Case for Cost of CO_2 Avoided	N/A	Ref 1a	N/A	Ref 2a	N/A	Ref 3a	Ref 2a	Ref 3a	Ref 3a	Ref 2a	Ref 1a

Case 1, a hybrid cycle consisting of a NGCC power plant and an atmospheric pressure MCFC equipped with cryogenic carbon capture technology, delivered the best economic performance of all CO_2 capture cases considered within this study review with a LCOE of 6.92 € cent/kWh. This represents a 23.9% decrease on the equivalent NGCC plus Cansolv PCC Reference Case 2b. The cost of CO_2 avoided is 33.5 €/t CO_2. This represents a 65.1% decrease on the equivalent NGCC plus Cansolv PCC Reference Case 2b.

A sensitivity study was carried out on Case 1. This showed that fuel cell investment cost had a significant impact on the LCOE performance compared to the other fuel cell parameters investigated:

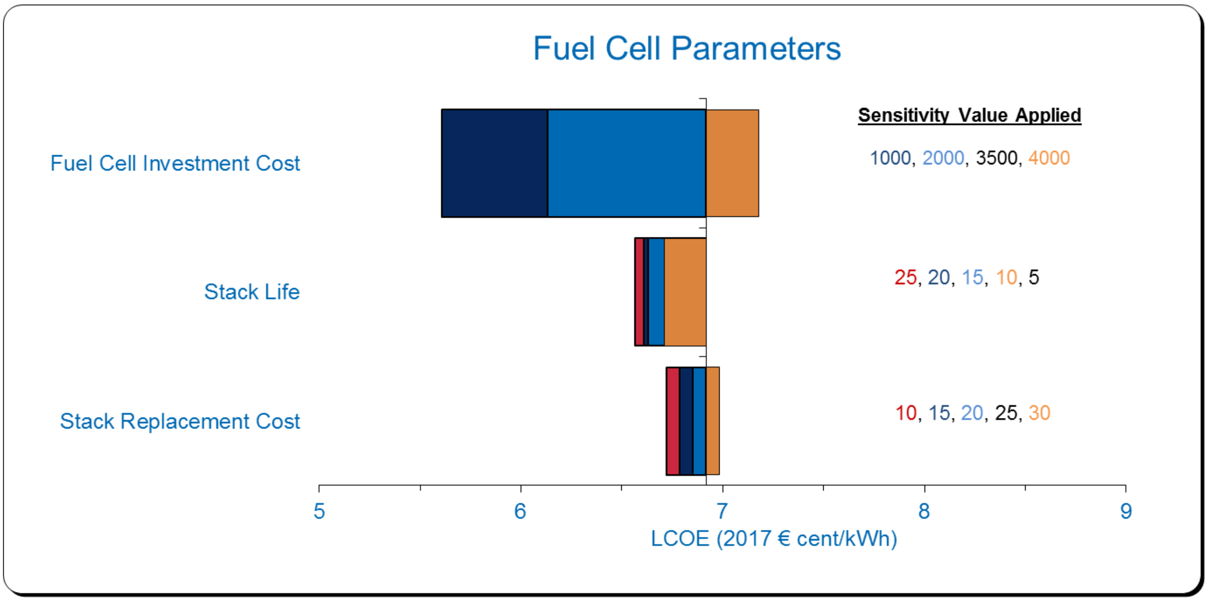


Figure 1: Case 1 Atmospheric NGFC-MCFC with CO_2 Capture: LCOE Sensitivity to FC Parameters

Reducing the FC cost from 3500 €/kW$_e$ to 1000 €/kW$_e$ (71% reduction in cost) results in a LCOE of 5.61 € cent/kWh, an 18.9% reduction when compared to the base case, Case 1. It should be noted that the 1000 €/kW$_e$ target was used based on targets set by the US Department of Energy for stationary combined heat and power fuel cell systems operating on natural gas (US DOE Fuel Cell Technologies, 2015). However, it is recognised that, for MCFCs a target of approximately 2000 €/kW$_e$ is more realistic (NREL, 2010).

A limitation with this study's approach is that base capital cost data was taken from a number of different studies collected in the literature review.

The capital cost data has been homogenised in terms of location, currency, year etc. However the raw data taken from the literature review was unchanged. Therefore a degree of variability is introduced from the different studies' base data and the accuracy and robustness of capital cost estimates they employ. This could potentially lead to varying results with comparatively optimistic or pessimistic cost assumptions. It should be noted that for all cases the base data for the FC capital cost was adjusted based on the findings of the literature review.

The results presented here, referenced to common base cases, are intended to provide a guide towards the most promising configurations and technologies within the described limits. For completeness the cost of CO_2 avoided has been included for each case as reported in the individual studies, it should be noted that these costs have not been adjusted or homogenised.

It is recommended that the cases are modelled to confirm performance as per literature with bottom up estimates made for each case using a single source of data. This will ensure that the results are developed using the same basis and methodology and are therefore directly comparable. However this was beyond the scope of this present study.

Key Conclusions

Based on the literature review and the techno economic analysis carried out in the present study, the conclusions were:

(i) Fuel Cell Technology

- Fuel Cell technologies with CO_2 capture is a promising alternative to conventional technologies and has a role to play in the next generation CCS.
- A number of studies can be found in the literature, focusing on SOFCs and MCFCs integrated in power cycles. However, it is evident that SOFC configurations have been the main focus recently. This may be due to the fact that SOFC technology has significant room for technical, hence economic, improvement. On the other hand, for MCFCs, the FC stack accounts for approximately 60% of the MCFC system costs (BOP, gas clean up, pre-reformer, water management, heat exchangers, control, inverter) (NREL, 2010). The same study concludes that no single issue could achieve significant cost reductions, but stack life time, power density and cost reduction of gas cleaning are the most important R&D areas to bring down overall system costs. However, even under the most optimistic circumstances, it is not likely that first costs for an MCFC power plant can be brought much below 2,000 US$/kW$_e$ (NREL, 2010).
- Several methods of CO_2 capture / separation can be employed together with FCs for further CO_2 purification:
  - Cryogenic CO_2 separation
  - Oxy-combustion
  - Oxidation
  - Condensation
  - Membrane Separation
  - Physical Absorption
- Of all the CO_2 capture methods, cryogenic CO_2 separation and oxy-combustion with condensation are the most widely considered.
- There are several factors that affect the performance of SOFC and MCFC, including:
  - Pressure
  - Temperature
  - Voltage
  - U_f
  - U_{Air} for SOFC and U_{CO2} for MCFC
- Optimum operational and plant parameters of the FC depend on individual design configurations. Some of the key plant areas and options for consideration are detailed in Appendix 2.

- Pressurised configurations showed significant potential with regard to the overall power plant performance. At a constant current density, increases in pressure result in an increase in voltage. However, with increasing pressure the voltage increases get smaller (Duan, et al., 2011). Therefore there is a diminishing return in pressure increases and eventually the increases in performance are outweighed by the significant costs and complexity of the system caused by pressurisation. For a given system the optimum operating pressure should be determined.

(ii) Techno-Economic Performance

- The techno-economic analysis showed that fuel cell systems plus CO_2 capture cases have the potential to be competitive with the current Cansolv based scrubbing system. However, limitations in the techno-economic analysis are noted. The cases taken from literature and updated in this present study are all based on simulated performance data and not on performance data of real plants. Therefore, validation of FC performance against the reference curves utilised would be required to determine the effects on the simulated overall system performance.
- Techno-economic Case 1, a hybrid cycle consisting of a NGCC power plant and an atmospheric pressure MCFC equipped with cryogenic carbon capture technology, gave the best economic performance of all CO_2 capture cases, with an LCOE of 6.92 € cent/kWh. This represents a 23.9% decrease on the equivalent NGCC plus Cansolv PCC Reference Case 2b.
- The cost of CO_2 avoided is 33.5 €/t CO_2 relative to NGCC Reference Case 2a for techno-economic Case 1. This represents a 65.1% decrease on the equivalent NGCC plus Cansolv PCC Reference Case 2b.
- The sensitivity analysis showed that the fuel cell investment cost was the key parameter affecting the LCOE for Case 1.
- Throughout the literature review, and as highlighted in the sensitivity study, FC investment cost is cited as the major barrier to commercialisation of FC plus CCS hybrid cycles. The key to a reduction in the cost for the FC is often linked to an increase in demand and hence volume of production, with the resultant savings driven through manufacturing efficiencies such as higher utilisation of equipment and labour. A major challenge is that for an increase in production to occur then the FC has to be at a price that makes the process economically attractive. In addition to cost reductions driven by R&D, time and scope-limited subsidies could be used as a driver for market introduction (Fuel Cells and Hydrogen Joint Undertaking, 2015).
- Based on the results of the study a reduction in FC investment cost from 3500 €/kW$_e$ to around 2000 €/kW$_e$ would be required to drive demand from an economics perspective assuming policy, fuel price etc. remain as is. In addition to savings anticipated through increases in production volumes, attempts to reach this FC cost could potentially be made through further improvements in material development where possible, improved leaner and automated manufacturing techniques, simplified designs and an engaged supply chain (Fuel Cells and Hydrogen Joint Undertaking, 2015).
- Enhancement of the current degradation rates in both SOFCs and MCFCs were identified as another key area to achieve a lower cost. However, the sensitivity study showed that, although important, the FC degradation rate effects were not as substantial as the effect of FC cost. Improvements in degradation rate should be targeted, but a greater emphasis should be placed on reducing FC costs to have the most significant impact on the LCOE, and hence economic viability of FC plus CCS hybrid cycles. The US Department of Energy's Solid State Energy Conversion Alliance (SECA) coordinates Federal efforts to facilitate development of a commercially

relevant and robust SOFC system. The key SECA objectives are an efficiency of greater than 60 percent, meeting a stack target cost of 175 US$/kW$_e$ and demonstrating lifetime performance degradation of less than 0.2 percent per 1,000 hours over a 40,000 hour lifetime. This emphasises the importance of both the cost and degradation improvements to achieve a commercial offering.

- Techno-economic Case 1 Atmospheric NGFC-MC with CO_2 Capture and Case 5 Atmospheric Hybrid-MCFC with CO_2 Capture both employ MCFCs utilising the flue gas from NG and PC processes respectively as a feedstock for the cathode. This is of particular interest, as it gives the potential for MCFC systems to retrofit NG and PC plants at lower efficiency penalty than current state-of-the-art amine scrubbing carbon capture, potentially with a lower CO_2 avoidance cost.

- Spinelli, et al.(2018) found that proposed MCFC retrofit configurations applied to PC and NG plants yielded a limited (<3 percentage points) decrease in efficiency, compared to ~6 to 8 percentage points efficiency penalty for amine scrubbing (NETL, 2015). Spinelli, et al. also found a CO_2 avoidance cost in the range 25 to 40 €/tonne. This shows good agreement with the results of the present techno-economic analysis, (33.5 €/tonne for Case 1). This is competitive with current state-of-the-art amine scrubbing carbon capture.

- FCs are commercially advantageous for power generation companies that are not tied to a particular type of fuel because FCs are flexible with regard to the feedstock. A company using SOFCs is prone to less risks of loss if one particular type of fuel becomes too expensive or even prohibited, since the fuel can be switched (with the appropriate processing and treatment steps).

(iii) Challenges

- Even though FCs show great potential to be a next generation CO_2 capture technology, there are still significant challenges that need to be addressed.

- As public policies begin to take shape, the concept of fuel cells combined with CO_2 capture could potentially become increasingly attractive since the FCs not only have higher electrical efficiencies in the power cycle, but also could potentially facilitate CO_2 capture at lower costs than conventional forms of fossil-fuel-based power generation. However, given their early stage of commercialisation, fuel cell systems will still require financial support mechanisms.

- An additional environmental benefit of FCs over conventional fossil power generation is the capability to minimise or even exclude the production of pollutants such as SO_X, NO_X and particulates. In moving to a clean economy the emphasis has been on CO_2 reduction, but given that every year almost 500,000 premature deaths have been linked to air pollution (European Environment Agency, 2018), there will potentially be a drive towards eliminating other air pollutants. This could potentially lead to the introduction of incentives / penalties that would contribute to an improved economic performance of FC systems.

- System manufacturing costs are expected to be reduced by increased mass manufacturing. The increasing average size of installations should fuel significant growth for the coming decade, but given price uncertainties, growth beyond that becomes less certain.

- Material and component costs are not expected to drop significantly with economies of scale. Further research is needed in that field in order to reduce FCs' installed costs and therefore make them competitive with other technology options.

Recommendations

- The results of this study show that FC with CCS hybrid cycles have the potential to be competitive with current state-of-the-art carbon capture technology but, as of yet there are no large-scale demonstrations which would be required to increase confidence in the technology and reduce the process risk. However, large projects are expected to be operational in the near future (Eisler, 2018) that will provide key performance data to enable further development of the next generation of carbon capture plants. This will increase confidence in the technology options and assist in overcoming the R&D challenges and barriers identified in the present study.
- The results presented here, referenced to common base cases, are intended to provide a guide towards the most promising configurations and technologies within the described limitations. It is recommended that the cases taken from literature are modelled to confirm performance as described, with bottom up estimates compiled for each case using a single source of data. This will ensure that the results are developed using the same basis and methodology and are therefore directly comparable.

CONTENTS

1. INTRODUCTION TO FUEL CELLS

1.1. Background

Fuel Cells (FCs) convert the chemical energy of a gaseous fuel directly into electricity and heat. Fuel is oxidised electrochemically, which leads to lower exergy losses compared to direct combustion. FCs consist of an electrolyte layer between an anode and a cathode, functioning as a membrane permeable for ions. The anode and cathode are connected in such a way that electrons generated at the anode flow to the cathode thus creating a closed circuit. In general, FCs are classified by the electrolyte material and their operating temperature, as seen in Figure 2. Low-temperature FCs include the alkaline FC (AFC), phosphoric acid FC (PAFC) and proton exchange membrane FC (PEMFC). Molten carbonate FCs (MCFC) and solid oxide FCs (SOFC) are high-temperature FCs. Due to the high temperature at which MCFCs and SOFCs operate, natural gas reformation and the subsequent shift reaction can be performed in the FC itself. MCFCs and SOFCs are most appropriate for stationary power production at scales ranging from a few hundred kW_e up to a few MW_e, due to their high electrical efficiencies and the ability for co-generation of electricity and heat (US DoE, 2004). The efficiency of SOFCs and MCFCs can be further increased by integration with gas turbines, as will be reported under this study.

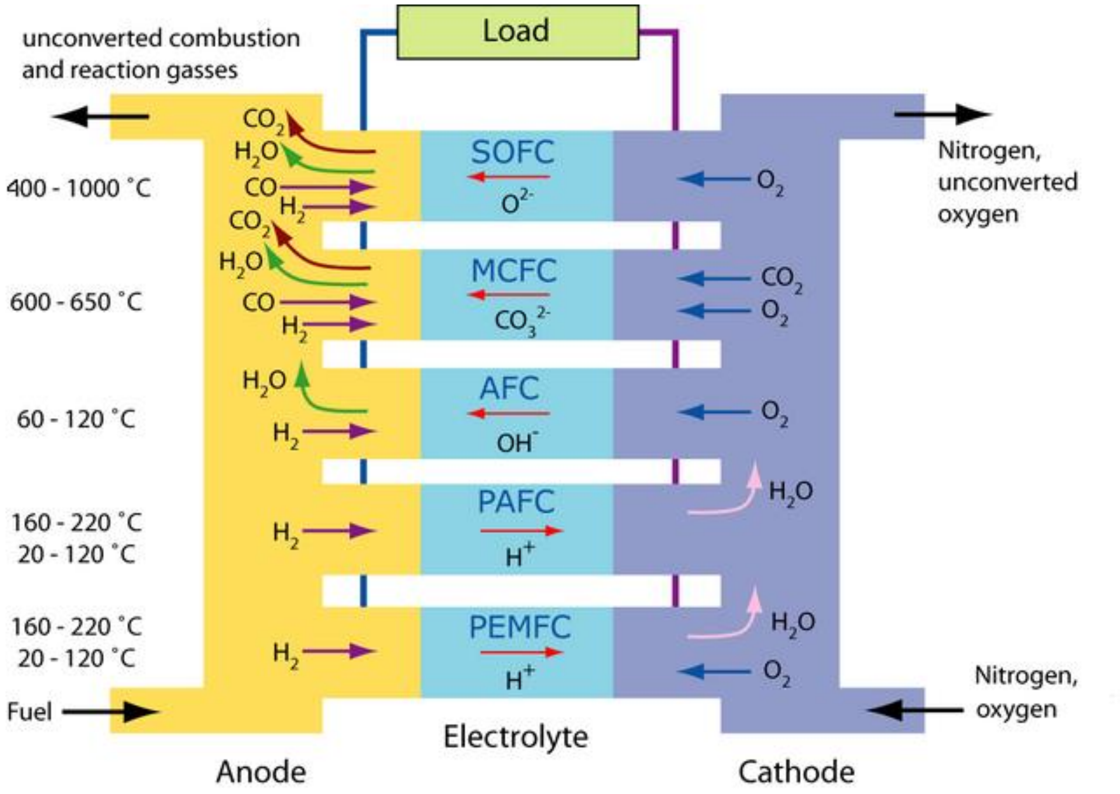


Figure 2: Superstructure of various FCs and their operating temperature (US DoE, 2004)

When FCs produce the entire electrical power output electrochemically, the apparent advantage is the elimination of the Carnot-efficiency limitation (Hassanzadeh & Mansouri, 2005). This fact, in parallel with environmental advantages (almost zero NO_x and particulate matter emissions), explain the increasing interest in fuel-cell systems integrated in combined cycles for power generation. Past literature has revealed the combustion process as a prime target for seeking improvements to existing, low energy-conversion efficiencies (Dunbar, et al., 1991). Present technology, with conventional fuel oxidation via the highly irreversible combustion process, utilises about 30% of the usable fuel energy. The reduction of such irreversibility's would obviously improve the efficiency of electricity production (Dunbar, et al., 1991).

In MCFC / SOFC configurations fuelled with natural gas or syngas, CO_2 capture can also be performed contributing to decarbonising the energy portfolio. CO_2 capture can be performed at various points, such as after the FC (in this report referred to as 'post-anode capture'). Alternatively, rather than utilising the SOFC / MCFC for reforming, H_2 can be produced by reforming/partial oxidation of natural gas or coal gasification upstream of the FC (Campanari, et al., 2013). CO_2 can be removed after syngas is shifted by means of physical solvents, membranes or adsorbents, in this report referred to as 'pre-anode CO_2 capture' strategy.

The use of external reforming also has potential for process improvements with techno-economic assessments of sorption-enhanced steam methane reforming (SE-SMR) and SE-SMR integrated with SOFC showing improved economic performance over conventional steam methane reforming with CO_2 Capture (Diglio, et al., 2017).

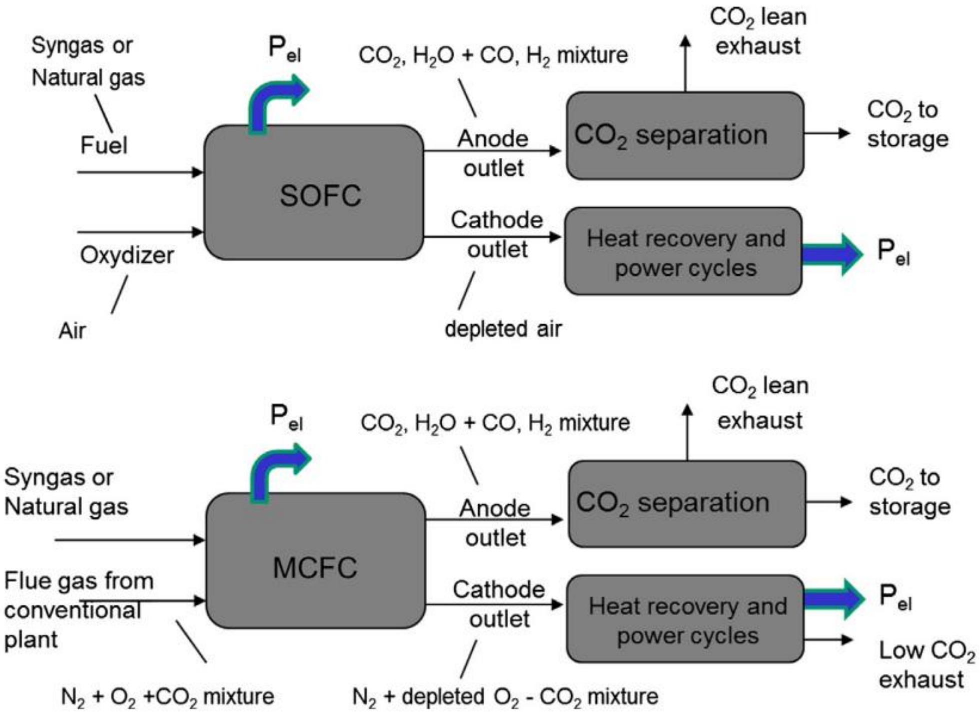


Figure 3: Two main options for CO_2 capture using FCs. On top, the FC oxidises a fuel taking oxygen from air and later separating CO_2 from the anode effluent. Below, the MCFC concentrates the CO_2 in flue gas from a conventional power plant from the cathode inlet to the anode outlet, while also oxidizing a portion of additional fuel. (IEA, 2008)

The concept of using FCs within large scale power cycles for CO_2 capture has been widely discussed, although most of the attention has been given to SOFCs (Li, et al., 2012) (Campanari, 2002).

Such systems generally operate with an approach which is similar to the 'oxy-fuel' concept, oxidising fuel with oxygen extracted from air while generating power, and releasing concentrated effluents at the anode outlet. This concept is shown at the top of Figure 3, where the anode exhaust is sent to the CO_2 separation island which captures CO_2 via chemical or physical separation technologies (post-anode capture) (IEA, 2008).

This approach for power cycles generally require integration with a custom-tailored gas turbine cycle, often operating at unconventional turbine inlet temperatures and pressure ratios, either using natural gas as a fuel or coal through Integrated Coal Gasification FC (IGFC) concepts. Since most fuel is oxidised in the FC to allow a high CO_2 capture efficiency, the FC generates the majority of the cycle power output (Verma, et al., 2006).

The alternative option offered by MCFCs is shown in the bottom block flow diagram in Figure 3, where the FC can operate ‘‘draining’’ CO_2 from the cathode inlet stream, receiving the flue gases of a conventional power plant. In this way the FC operates with a post-combustion approach, although also oxidising a minor portion of additional fuel with the same 'oxy-fuel' features discussed above.

In terms of fuel cell performance the ideal standard potential of an H_2 / O_2 fuel cell is 1.229 V with liquid water as the product and 1.18 V with gaseous water as the product. Figure 4 shows the influence of temperature, particularly pronounced for high temperature fuel cells such as MCFC and SOFC, on the standard potential. Hence the ideal and actual performance of a fuel cell is quite different.

Figure 5 shows the ideal and actual response of a fuel cell Electrical energy is obtained from a fuel cell when a current is drawn, but the actual cell potential is lowered from its equilibrium potential because of irreversible losses due to various reasons. Several factors contribute to the irreversible losses in a practical fuel cell. The losses, which are generally called polarization or over potential, originate primarily from activation polarization, ohmic polarization, and gas concentration polarization (Pilatowsky, et al., 2011).

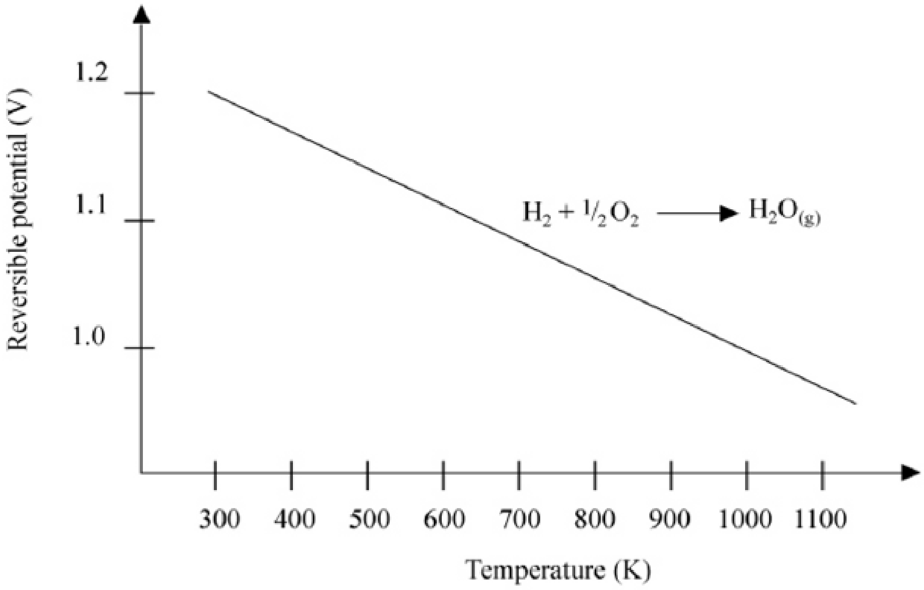


Figure 4: The influence of temperature on the standard potential of an H2 / O2 fuel cell (Pilatowsky, et al., 2011)

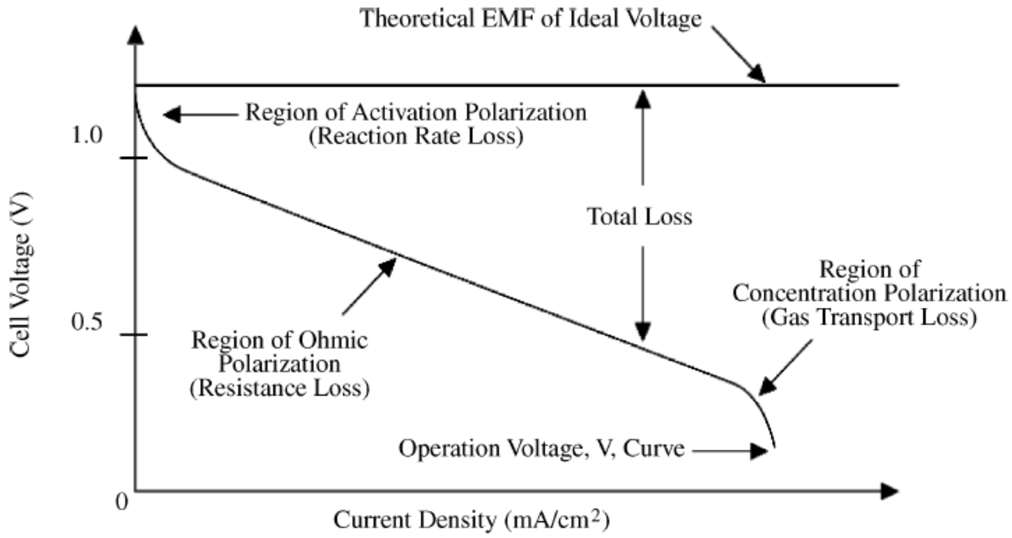


Figure 5: Ideal and actual performance of a fuel cell with respect to the potential current response (Pilatowsky, et al., 2011)

1.2. Study Objectives

The purpose of this Review Report is to:-

- Conduct a review and discuss various capture options in an attempt to identify the optimum Fuel Cell power cycle configuration integrated with CO_2 capture in terms of reported Levelised Cost of Electricity (LCOE), overall plant efficiency and complexity.
- Review key factors affecting the performance of the power cycle in an attempt to identify bottlenecks, barriers and challenges for the commercialisation of such CO_2 capture concepts and next steps required for future research will be suggested.

2. SOLID OXIDE FUEL CELLS

2.1. Background of SOFC technology

Solid Oxide FCs (SOFCs) are highly efficient electrochemical devices that convert the chemical energy of a fuel into electricity avoiding the necessity of direct combustion. The SOFC cell is constructed with two porous electrodes (anode and cathode) that surround an electrolyte, as seen in Figure 6.

Air flows along the cathode, while fuel is introduced into the anode. When an oxygen molecule contacts the cathode/electrolyte interface, it acquires electrons from the cathode. The oxygen ions diffuse into the electrolyte material and migrate to the other side of the cell where these contact the anode. The oxygen ions meet the fuel at the anode/electrolyte interface and react catalytically, producing water, carbon dioxide, heat, and electrons. The electrons transport through the external circuit, providing electrical energy.

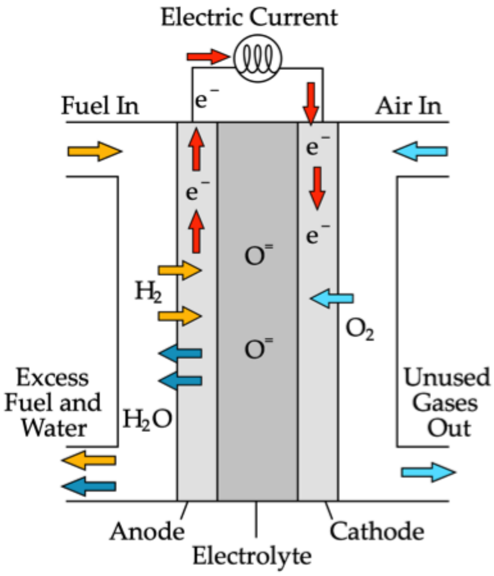


Figure 6: Configuration of the Solid Oxide FC (SOFC) (US DoE, 2004)

Solid oxide FCs (SOFC) allow conversion of a wide range of fuels, including various hydrocarbon fuels (syngas, natural gas, hydrogen, and other hydrocarbon fuels such as gasoline or diesel). SOFCs operate at relatively high temperatures between 700-1000°C, depending on the material used for the manufacture of the cell stack. SOFCs can operate both at atmospheric and pressurised conditions, however, their performance and efficiency increase when operated under pressure. Because of their high temperature and operating pressure, SOFCs are considered ideal for integration in power generation systems for replacing the thermodynamically constrained gas turbine (GT), however, it should be noted that SOFCs do not have the same load following capability as a GT. These operating conditions allow for highly efficient fuel to power conversion, internal reforming, and high quality by-product heat for cogeneration or for use in a bottoming cycle. As a result, in addition to the electricity produced electrochemically by the SOFC, the outlet gas streams from the SOFC stack can be expanded in a GT to produce additional power.

Steam can also be generated utilising the heat of the exhaust gas to power a steam Rankine Cycle. Indeed, both simple-cycle and hybrid SOFC systems have demonstrated among the highest efficiencies of any power generation system, combined with minimal air pollutant emissions and low greenhouse gas emissions.

Additionally, the outlet gas streams can be easily integrated with CO_2 capture technologies to yield clean power. These capabilities have made SOFC an attractive emerging technology for stationary power generation in the 2 kW_e to 100s MW_e capacity range. However, a major remaining challenge is the comparable scale up of the SOFC unit for large scale power plants due to cost and material constraints.

2.2. SOFC systems integrated in power plant cycles

2.2.1. First-Generation SOFC Systems

First-generation SOFC systems are the most promising option for SOFC commercialisation at large scale (100 MW_e or greater) in the near future (Adams, et al., 2013).

Several process configurations are possible and the various design options can be seen as a superstructure in Figure 7. While many process configurations are feasible, these generally follow the same pattern and can be divided in five main sub-islands:

- fuel preparation island;
- SOFC power generation island;
- fuel completion island;
- heat recovery island; and
- CO_2 Capture/Purification and Sequestration system island.

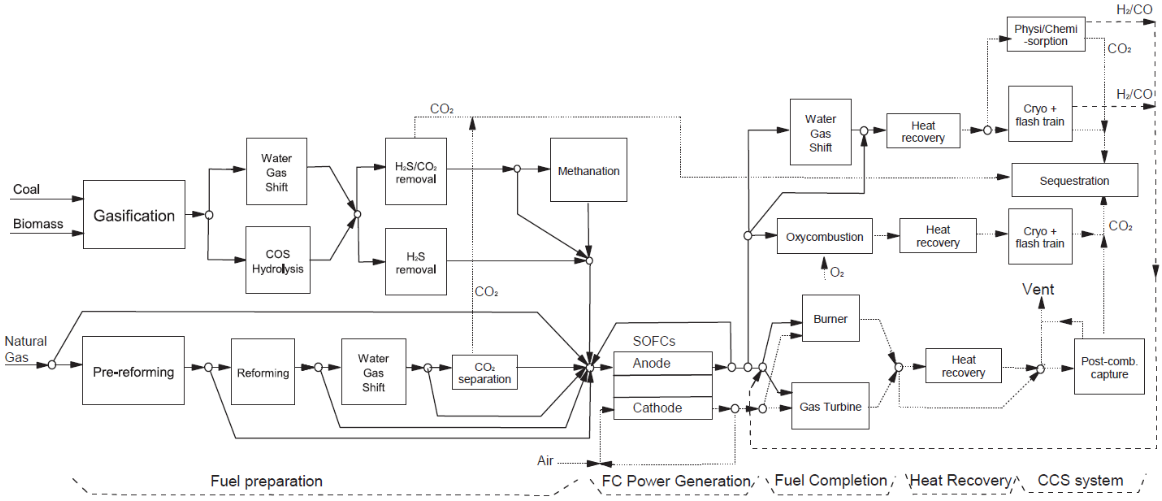


Figure 7: Superstructure of the various designs possible for SOFC integrated with CO_2 capture (Adams, et al., 2013)

A. Fuel Preparation Island

In the fuel preparation island, the raw fuel (such as natural gas, coal or biomass) is prepared prior to entering the SOFC unit.

For the case of coal and biomass, a gasification process is required in order to produce syngas (a mixture of H_2, CH_4, CO and CO_2). This is followed by cleaning steps to remove HCl, formed during gasification of chlorine containing coals or biomass, which can seriously affect the efficiency of the SOFC causing significant current density losses at concentrations as low as 160 ppm (Trembly, et al., 2007), sulphur-containing compounds (depending on the SOFC materials used, H_2S causes adverse effects at concentrations of 1−120 ppm (Marquez, et al., 2007)), siloxanes (a common biogas contaminant that causes silica to be deposited on the interconnector and anode current collector. Even trace contamination of the fuel feed at ppb level can affect the SOFC Ni anode leading to fast degradation (Madi, et al., 2015)), higher hydrocarbons (carbon may diffuse into the nickel structure to a point where physical properties are altered and/or nickel metal dusting occurs, thereby diminishing the anode reaction activity, corrupting microstructure, and disrupting the conductive pathways (Hackett, et al., 2015)) ammonia, and mercury.

An optional water-gas-shift (WGS) reaction may be used to upgrade the hydrogen content of the syngas by converting H_2O (steam) and CO to H_2 and CO_2:

$$H_2O + CO \rightarrow H_2 + CO_2 \qquad (2\text{-}1)$$

High-efficiency systems have been proposed both with and without a WGS reaction step; previous studies have reported certain power losses or degradation effects associated to carbon deposition due to the lack of a WGS step (Kee, et al., 2008), however, un-shifted coal-derived syngas has been used successfully in other experiments (Lim, et al., 2008). As a result, the decision on a WGS step should consider other parameters such as trade-offs associated with the costs of coke-resistant anode materials, the actual gains in SOFC stack lifetime achieved by avoiding CO feeds and the costs of a WGS unit. The second reason to include a WGS is that it might be employed specifically for the purposes of H_2 generation and CO_2 capture. When the syngas is fully shifted, the fuel stream becomes a mixture of H_2 and CO_2. Before entering the anode of the SOFC, the CO_2 can be captured via absorption, for example using solvents, adsorption or membrane technologies, leaving a stream of mostly H_2 to use for the cell. This enables upstream, pre-anode CO_2 capture without requiring CO_2 separation downstream.

An optional methanation process can be utilised with the aim of increasing the methane content in the fuel gas, reducing the air flow rate for SOFC cooling and improving the energy conversion efficiency of the integrated power cycle (Romano, et al., 2011).

For the case of natural gas, the steam reforming reaction can be used to convert some or all of the hydrocarbons in the gas feed into H_2 and CO, following the reaction below:

$$C_nH_m + nH_2O \rightarrow nCO + (n + m/2)H_2 \qquad (2\text{-}2)$$

The steam reforming reaction is highly endothermic. Therefore, a pre-reforming reactor (reforming of higher hydrocarbons) and/or a reforming reactor (reforming of methane and others) may be used for heating purposes and thermal management. This process will be referred to as ‘external reforming’ in this report. Since SOFCs operate at high temperatures, as discussed above, it is a common technique to place reforming zones adjacent to the SOFC anode so that heat is transferred from the anode to the reformer by radiation (Riensche, et al., 2000). If natural gas reforming does not take place before entering the anode, reforming can still occur inside the anode itself, which is known as “internal reforming”. The result is similar to the case of including/excluding a WGS, since high efficiencies are reported for both external and internal reforming, commercial implementation will depend on the trade-off between capital equipment costs, operating costs, efficiency, and SOFC lifetime (Braun, et al., 2012).

The anode exhaust gas consists of unspent fuel (H_2 and CO) with the quantity dependent on the fuel cell’s fuel utilisation (U_f) and the by-products of the electrochemical reaction (H_2O and CO_2).

B. SOFC Power Generation Island

Design decisions within the power generation island focus on the use and extent of anode/cathode gas recycling and the amount of fuel utilisation (U_f), which is the percentage of fuel that is electrochemically oxidised in the cells. Setting the U_f factor too high or too low will result in diminishing returns on the performance of the SOFC. As such, 100% utilisation has not been studied and a 70-90% is more common, with most of the cases investigated in this report using a U_f between 75% and 85%.

Regarding gas recycling, like many typical chemical processes, the use of anode and/ or cathode recycling can improve the performance of the cell (Cheddie & Murray, 2010) (Lee, et al., 2011). However, this comes with certain complexity trade-offs, such as the use of recycle compressors and additional piping. However, other benefits can also be realised such as improving excess heat recovery and eliminating the need for external steam production for the anode, which contributes to an overall reduction in costs of electricity production. This has been shown through modelling for 2 kW_e (Braun, 2010), 200 kW_e (Riensche, et al., 1998), and 12,000 kW_e scales (Cheddie & Murray, 2010), although it may be optimal to keep the amount of anode recycling to a relatively low amount (Lee, et al., 2011).

C. Fuel completion, Heat Recovery and CCS System Islands

Since the U_f factor varies between 75% and 85% the anode exhaust will contain some amount of unspent fuel. Instead of venting this fuel to the atmosphere, it is desirable to use this remaining energy in some way. For low-pressure and high U_f systems (>70%), a simple option is to oxidise the fuel using air in a burner, with the cathode exhaust providing enough unspent O_2 for this process. This way a waste stream is created which may be used for downstream heat recovery – upstream preheating, steam generation, or hot water production. Less common options could include absorption refrigeration cycles, thermo photovoltaic power generation, Cheng cycles, and humid air turbine cycles that were reviewed by (Zhang, et al., 2010). If steam is generated, it can be used for additional electricity production in a bottoming cycle (e.g. Rankine or Brayton cycle) to further increase the power output of the plant.

Instead of burning with air or cathode exhaust, it is also possible to oxidise the fuel using a high-purity O_2 stream. This will also generate a hot anode exhaust stream for heat recovery, except that there will be no nitrogen, leaving only CO_2 and H_2O. If this stream is not mixed with the cathode exhaust, the CO_2 and H_2O can be relatively easily separated after heat recovery by condensing out the H_2O, leaving a high-purity CO_2 stream suitable for pipeline transport (Adams & Barton, 2010). However, in the case of low pressure SOFC systems, the CO_2 captured is approximately at atmospheric pressure, which means that the energy required to pressurise the CO_2 up to the supercritical pressure (150 bar) required for pipeline transport will be quite large. Second-generation (high-pressure) SOFC systems that will be discussed below are likely to be a better option for this purpose since the CO_2 can be captured at higher pressures. In a third alternative, if the U_f is <85%, the amount of unspent fuel in the anode exhaust may be large enough to permit combustion in a gas combustion turbine for extra power production. These hybrid SOFC / GT systems have other unique advantages and are described in the next section.

From an efficiency point of view, however, it is generally preferred to have a higher U_f factor and produce less electricity from utilising waste heat, since the thermodynamic efficiency of a SOFC unit is typically higher than the Carnot cycle, as discussed previously. There are various studies that have focused on sensitivity analyses of U_f and how it affects the overall system efficiency. It has been reported that the total system efficiency increases by roughly 1.5 percentage points (HHV) for every 5% increase of the U_f factor for one particular system configuration (Adams & Barton, 2010).

2.2.2. Second-Generation SOFC Systems with CCS

Second-generation SOFC systems are high-pressure SOFCs with separate streams for the anode and cathode exhausts. This promotes the use of an SOFC system which captures and compresses CO_2 at significantly reduced costs and minimum complexity via ‘pre-anode’ and/or ‘post-anode’ capture, as seen in Figure 7.

In the pre-anode process, syngas is generated at high pressure via high pressure coal gasification or by high pressure natural gas reforming. Similar to the first generation SOFC systems, the syngas can be optionally shifted using the WGS reaction, creating a stream of steam, H_2, and CO_2. Up to about 90% of the CO_2 can then be recovered from the syngas (or shifted syngas) using absorption technologies. This strategy is similar to the one used for Integrated Gasification Combined Cycle (IGCC) pre-combustion CO_2 capture process.

Alternatively, the “post-anode” capture strategy can be applied in which CO_2 is separated from H_2O via H_2O condensation (or via cooling, knock-out and additional drying). Post-anode CO_2

capture can effectively result in a 100% CO_2 removal (Adams & Barton, 2010). A separation system which uses condensation followed by a cascade of flash drums, with final drying step can be used to produce CO_2 at high enough purity for pipeline transport at the SOFC anode exhaust pressure, which can result in total system efficiencies of 46%, HHV, net for coal-based systems and 73%, HHV, net for natural gas fuels (Adams & Barton, 2010). Other condenser-based strategies have been reported with similar efficiencies (44−59% for coal, HHV; up to 72% for natural gas, HHV), as summarised in Table A of Appendix 1 and will be discussed later in Section 2.4 of the report.

Since the U_f of the SOFC is below 100%, the fuel completion island usually includes an oxidation process where the remaining fuel is oxidised in high-purity O_2 (Figure 7). Oxidising all the H_2 and CO from the SOFC anode exhaust with air would result in a too high dilution of the stream with nitrogen. Lygre et al. (Lygre, et al., 2001) studied oxidation with pure oxygen, but this resulted in significant additional costs and energy consumption if oxygen is not available. In another concept developed by Shell it was chosen to use an oxygen-conducting membrane reactor (OCM-reactor) placed after the SOFC (Adams, et al., 2013). The anode off-gas is fed to one side of the membrane while the cathode off-gas is fed to the other side. The membrane is selective to oxygen, which permeates from the cathode off-gas stream to the anode off-gas. In the membrane unit, H_2 and CO are oxidised. The retentate of the membrane unit consists of CO_2 and water. The water can be removed using the aforementioned conventional techniques.

For post-anode CO_2 recovery, it is worth mentioning that there is significant potential to capture liquid water in large quantities as well as CO_2. This water is at high purity which, after degassing the dissolved CO_2, can be re-used elsewhere to reduce the raw water consumption of the overall process. Depending on the cooling strategy of the system, recycling the water obtained by CCS can reduce the plant's water consumption by 30−35%, (Adams & Barton, 2010). Overall, the SOFC plant with CCS would use 24−57% less water compared to pulverised coal power plants, depending on the cooling process of the conventional power plant (once-through cooling or wet cooling towers).

In the case of using alternative (i.e. waterless) cooling strategies, such as air-cooling, the SOFC plant with post-anode CO_2 capture could actually result in a net production of high-purity water as a by-product of the order of 1 billion litres per year for a 712 MW_e plant (including approximately 50 MW_e of parasitic load for the air-cooling) (Adams, et al., 2013). Natural gas fuelled SOFC power plants with post anode CO_2 capture could show even better results; up to 1.5 billion litres per year of water can be produced for a plant of similar size (700 MW_e) (Adams & Barton, 2010). The significant reduction in water consumption (or even the ability to generate high purity water) can have significant long-term commercial benefits since access to fresh water in many parts of the world is an issue of increasing significance. As access to clean water becomes more expensive or difficult, the SOFC-based process could become increasingly attractive.

2.2.3. SOFC/GT Hybrid Systems

A hybrid FC – gas turbine (GT) system is arranged in such a way that either the FC is bottoming the turbine or the turbine is bottoming the FC. Both configurations aim at maximising the electrical output – and consequently the efficiency – of the overall system. Another benefit of this hybridisation is the elimination of parasitic electric loads on the power system from cathode air blower requirements. Even in natural gas fuelled systems with internal reforming, excess cathode airflow is required for thermal management as well as oxidant flow.

While it is possible to use SOFCs as the bottoming unit, it is generally not preferred. Instead, molten carbonate FCs (MCFC) are preferable for such a configuration, since MCFCs require CO_2 and O_2 in the cathode side, which ties up well with the turbine outlet. Such configurations are commercially available for natural gas fuelled systems (Ghezel-Ayagh, et al., 2005). In the case of the SOFC, most configurations make use of a direct fired configuration with the FC operated under pressure, topping the gas turbine. This configuration results in the highest theoretical electric efficiency and the highest power density and consequently the lowest overall system cost. Both Mitsubishi Heavy Industries, with a segmented tubular SOFC (Gengo, et al., 2005) and Rolls Royce (LG) FC Systems, with a micro tubular SOFC are currently pursuing this type of hybrid power system (Nichols, et al., 2008).

A significant drawback of the pressurised SOFC / GT hybrid is its complexity in terms of component integration and system controls. More examples will be analysed in the next section and can also be seen in Table A of Appendix 1.

Duan et al. (Duan, et al., 2011) employed an exergy analysis method to investigate the effects of different operating parameters on a modelled zero-CO_2 emission SOFC hybrid power system.

The fuel utilisation factor has a significant effect on the cell voltage and efficiency. Figure 8 shows that with an increase of fuel utilisation factor from 0.78 to 0.96, the current density will increase which will result in the decrease of cell voltage.

At lower values of fuel utilisation factor the cell voltage change is not significant for increases in fuel utilisation factor, so the system output exergy increases. For higher fuel utilisation factor, the change of amount of the cell voltage is bigger than that of the current density, as a result, the system exergy efficiency reduces.

Fuel utilisation factor also has a significant effect on the anode exhaust stream composition with a higher fuel utilisation factor resulting in less H_2 and CO in the anode exhaust (Duan, et al., 2011).

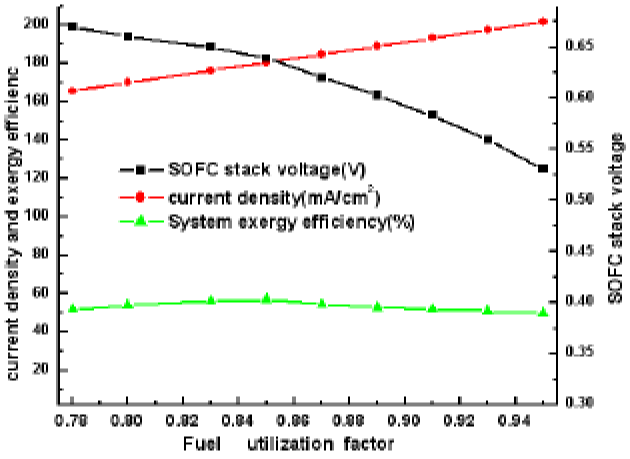


Figure 8: The effect of fuel utilisation factor on system performance

2.2.4. SOFC and effect of flue gas contaminants

SRI International (SRI International, 2008) performed a study to determine the sensitivity of the performance of solid oxide fuel cells to trace level contaminants present in a coal-derived gas stream in the temperature range 700°C to 900°C.In short-term accelerated tests, eight different potential contaminants were tested that can be present in the anode feed from a coal gasifier. Table 2 summarises the test conditions and degradation in power density after 100 hours of exposure.

Table 2: Degradation of the SOFC Coupons on Exposure to Coal Contaminants after 100h (SRI International, 2008)

Contaminant	Exposure Level (ppm)	750°C	800°C	850°C
As (As_2(g))	10	10	Cell Failure	
P (HPO_2(g))	40	7.5	10	
Cd(g)	5		<1	8
CH_3Cl(g)	40		<1	4
HCl(g)	40	<1	<1	
Hg(g)	7	<1	<1	
Hg(g)	0.18	<1	<1	<1
Sb (SbO(g))	8	<1	<1	1
Zn(g)	10		<1	<1

The results of the accelerated tests indicated that AsH_3, PH_3 and CH_3Cl will have a significant effect on the performance of the SOFC anodes. These contaminants at trace levels together with 1 ppm of H_2S were tested in long term tests.

H_2S at ppm levels was found to cause an immediate performance drop during the first few hours but no long term effect was observed. When the anode fuel gas contains multiple contaminants including H_2S and AsH_3 or PH_3 at 1 ppm level the cell power density decreased more rapidly than with one contaminant alone.

Chloride containing gases such as CH_3Cl and HCl at ppm level can also cause immediate but slight cell power density decrease. However, chloride containing species may reduce the degradation effect of AsH_3 and PH_3 significantly. So its overall effect may be beneficial.

The poisoning effect of S, Cl, As, and P containing species can be mainly differentiated into two categories: surface effect, or bulk effect. S and Cl adsorb on the anode surface, blocking the active catalyst sites for fuel adsorption and reaction, and their effect is recoverable. Whereas As and P can react with the anode, forming irreversible compounds, and therefore, the effect is not recoverable (SRI International, 2008).

2.3. Selection of the SOFC control strategy

One important factor that affects the performance and the lifetime costs of a FC system is the degradation rate of the cell stack. The performance of a FC tends to decrease over its lifetime due to degradation phenomena that occur in the electrodes, the electrolyte, the sealing materials and the metallic interconnects (Wuilleim, 2009). Endurance of the stack is critical for the commercialisation of the SOFC technology.

It was demonstrated that the evolution of the degradation, and consequently the reduction of the stack lifetime, is strongly affected by the SOFC control strategy (Thijssen, 2009), which can be one of the following two:

- constant current density operation (constant j);
- constant voltage operation (constant V)

Generally, the degradation shifts the operating point from the nominal condition to less performing ones over time. This issue can be simplified assuming that degradation yields an increase of the overall stack resistance, generally defined by the area stack resistance (ASR), and that the operating point is moved toward increased polarisation levels (Curletti, et al., 2015). Figure 9 shows the time evolution of the operating conditions with respect to the different control strategies (j = constant and V = constant). This figure refers to a simple linear polarisation model with a fixed ASR that does not depend on the current density value.

If, in a given period, the ASR changes from ASR_1 to ASR_4 (with $ASR_4 > ASR_1$), the resulting point is shifted vertically in case of constant current density operation strategy and horizontally in the constant voltage mode.

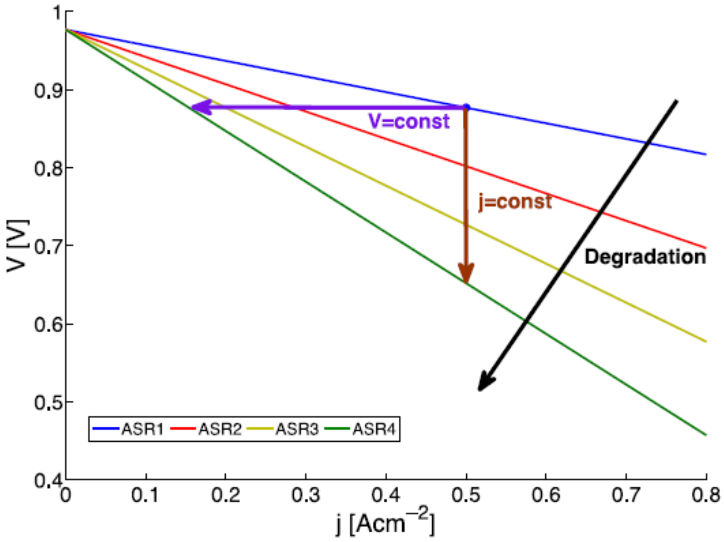


Figure 9: Time evolution of the operating conditions in respect to the different control strategies (Curletti, et al., 2015)

Under constant current operation, the ASR decreases very rapidly and so too does the system performance. In this mode, in order to ensure a continuous productivity during time, it is necessary to oversize the whole plant and to replace the stacks of the FCs when its capacity is lower than 90% of the initial capacity.

Under constant voltage operation, the degradation of the ASR is slower than under the constant current operation mode, and a strategy based on the activation of spare capacity can be adopted to compensate for any performance losses. Again, the spare capacity requires plant oversizing but the replacement of the stack can be done when the capacity is lower than 30-50% of the initial one (Curletti, et al., 2015).

Another key aspect of SOFC operation and control is thermal management. SOFC stack life is directly linked to stack performance degradation which depends partly on the structure and materials but also on the operating conditions of the stack.

Rapid temperature changes can cause mechanical stresses due to the differences in thermal expansion of the ceramic and metal components. This can cause significant issues in SOFC operation and performance particularly if there is a large degree of thermal cycling. This is particularly a cause for concern during transient operation such as start-up, shut-down and load following (Barelli, et al., 2016).

2.4. SOFC for CO_2 capture

2.4.1. Pre-anode CO_2 capture

Liese (Liese, 2010) studied four different cases of SOFC Integrated Coal Gasification Fuel Cell (IGFC) power plants, two of which used a selexol CO_2 capture process before the SOFC unit (Pre-Anode CO_2 capture) – case SO7 and SO8 in Table A of Appendix 1.

Liese Case SO7 : SOFC IGFC + Selexol CO_2 Capture + WGS

This study showed that pre-anode capture can have a higher efficiency than post-anode capture if the CO_2 is recovered at low pressure. The configuration of case SO7 includes cathode exhaust recycling by using a blower which preheats the inlet air, thus removing the need for a recuperator, as seen in Figure 10.

Recycling will have a negative effect on SOFC performance (i.e. a lower Nernst voltage) by lowering the oxygen concentration in the cathode. However, using cathode recycle, a higher turbine exhaust temperature can be achieved (600°C), high enough to utilise a heat recovery steam generator (HRSG). In this case CO_2 capture is achieved in the Fuel Conditioning section, prior to entering the anode, by using a Selexol process for simultaneous H_2S and CO_2 removal. High pressure (HP) steam from the syngas cooler is added, resulting in a steam to CO ratio of 1.8 for the WGS reactor. The exothermic WGS reaction is kept at a constant temperature with the low pressure (LP) steam generator. The end CO_2 capture rate is 87%, instead of 95%. This is due to the methane content of the syngas. To avoid this drop in CO_2 capture some further gas recycling or a polishing step would be needed. In this case, since a hydrogen-rich syngas is produced, carbon dusting is not a concern and thus preheating to 650°C for the anode inlet is done with heat exchange from the anode exhaust. Then the anode exhaust is directed to the GT combustor.

This configuration resulted in a 52%, HHV net efficiency for a 621MW$_e$ pressurised (9 bar) SOFC system. For the pre-anode systems studied by Liese et al., the parasitic load of pre-anode CO_2 capture process was offset by improved FC performance since the fuel mixture had a high partial pressure of H_2 after the CO_2 had been removed.

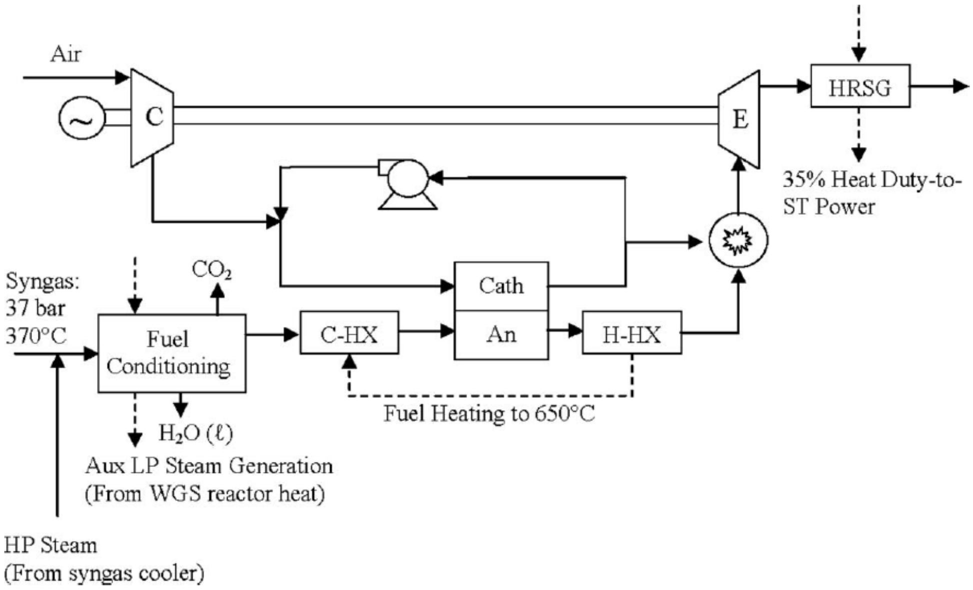


Figure 10: Schematic diagram of case SO7 (Liese, 2010)

Liese Case SO8 : SOFC IGFC + Selexol CO_2 Capture

In the case SO8 (Liese, 2010), the cathode configuration remains the same as in the case SO7, and CO_2 is still removed upstream to the anode. However, in this case, there is no HP steam addition or WGS reactor in the Fuel Conditioning block.

As a result, CO remains part of the syngas composition, with WGS taking place internally in the SOFC. Anode recycle is employed for preheating and to prevent carbon deposition. Addition of supplementary low pressure steam and a post-anode WGS reaction are also employed to prevent carbon deposition. A small fraction of the anode exhaust goes to the GT combustor, and the remainder is expanded to atmospheric pressure. However, the expanded fuel is not oxidised. The heat is used to produce low pressure steam. After the WGS reactor, water is condensed leaving a gas mixture of H_2, CO, and CO_2, along with N_2 and Ar which is recycled back to the Fuel Conditioning block upstream to the CO_2 removal process. A schematic diagram of the case SO8 can be seen in Figure 11.

The motivation for the second fuel recycle in case SO8 is due to two reasons. First, the high pressure steam addition needed in case SO7 for the WGS results in a significant energy penalty, so eliminating this step would be beneficial. Secondly, it is useful to utilise as much H_2 in the SOFC as possible, since the SOFC is more efficient than the GT unit, as discussed previously. However, the gas composition can have a limiting effect, i.e., too much CO_2 will limit the WGS reaction – and thus the production of H_2 from CO – and too little H_2 (or too much H_2O) will affect the SOFC electrochemical performance. The addition of the second recycle effectively allows for increased U_f, since the SOFC performance is not affected. As a result, more CO is shifted to H_2 and the exothermic reaction takes place in the SOFC, thus, producing a high grade waste heat.

Case SO8 did not produce very different results in terms of efficiency – 53%, HHV overall system efficiency – even though the fuel recycle resulted in a much higher SOFC power output, namely 753 MW_e.

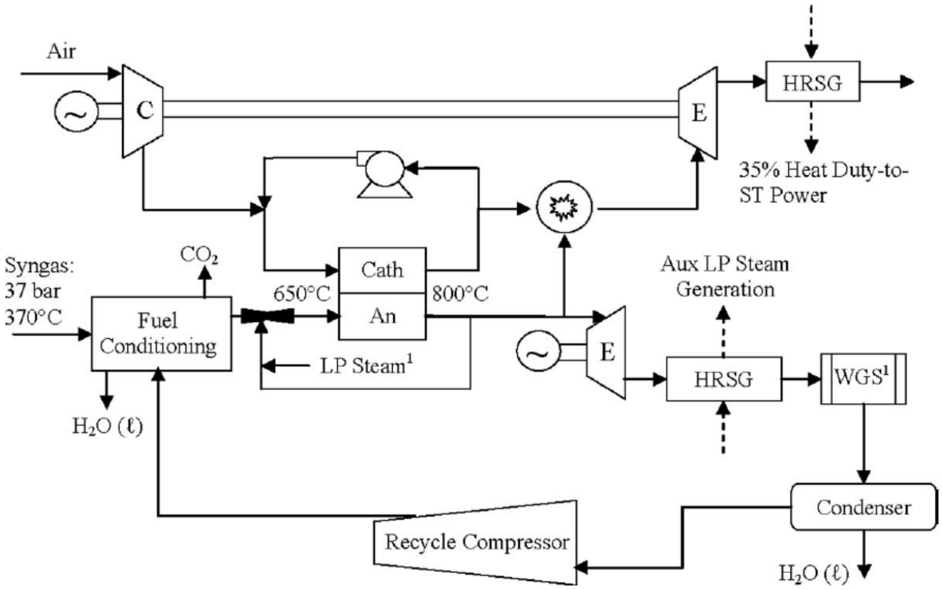


Figure 11: Schematic diagram of case SO8 (Liese, 2010)

Siefert et al. (Siefert & Litster, 2013) studied the pre-anode CO_2 capture in an IGFC with a catalytic gasifier and a pressurised SOFC (5 bar) using a Chemical Looping Combustion (CLC) approach.

Contrary to Liese's approach, the use of commercially available processes for the CO_2 capture such as Selexol (UOP LLC) or Rectisol (Linde AG and Lurgi AG) was avoided, as these are required to lower the temperature to below the dew point of the syngas. Instead, Siefert et al. modelled a chemical capture process that leaves the temperature of the syngas close to the inlet temperature of the SOFC. After leaving the catalytic gasifier, the methane rich syngas goes through an expander to drop the pressure of the gas to the pressure of the SOFC. The gas is then directed to a reactor filled with magnesium and calcium oxide (MgO, CaO) in order to capture CO_2 as well as any remaining H_2S and COS in the gas stream (Blamey, et al., 2010). CO_2 capture takes place at approximately 750°C, depending on the pressure after the expander. The CO_2 is regenerated from the dolomite ($MgCO_3$, $CaCO_3$) using hot exhaust gases from the SOFC at a temperature of 1000°C and a pressure of 0.1 MPa. Then, the CO_2 is cooled, dried, and then compressed to a pressure of 15 MPa for transport and storage. Siefert et al. reported an efficiency of 58%, HHV, net with a levelised cost of electricity (LCOE) of 52 US$/MWh which included revenues from EOR sequestration and 60 US$/MWh with revenues from saline sequestration.

Another study that focused on CLC was proposed by Aghaie et al. (Aghaie, et al., 2016), who presented a novel integrated system combining biomass gasification, pre- and post- CLC pressurised SOFC system (12 bar) and a steam power cycle.

In this process part of syngas exiting the gasifier enters the chemical looping reactor and another part enters the reformer which operates at 950°C and 22 bar. Syngas from the gasifier (1400C) is used to pre-heat the air prior to entering the SOFC cathode. Syngas from the reformer enters the anode which operates at 900°C and 12 bar. Outlet gases from the anode enter the chemical looping reactors which play the role of a combustion chamber, whereas outlet gases from the cathode are used for power production in the gas turbine. Outlet hot exhaust gases from the turbine are directed to the HRSG. This process can be seen in Figure 12. The net efficiency of the system reaches 55.8%, HHV, with 100% carbon dioxide capture, however, the Technology Readiness Levels (TRLs) of such an integrated system is relatively low (~3-5).

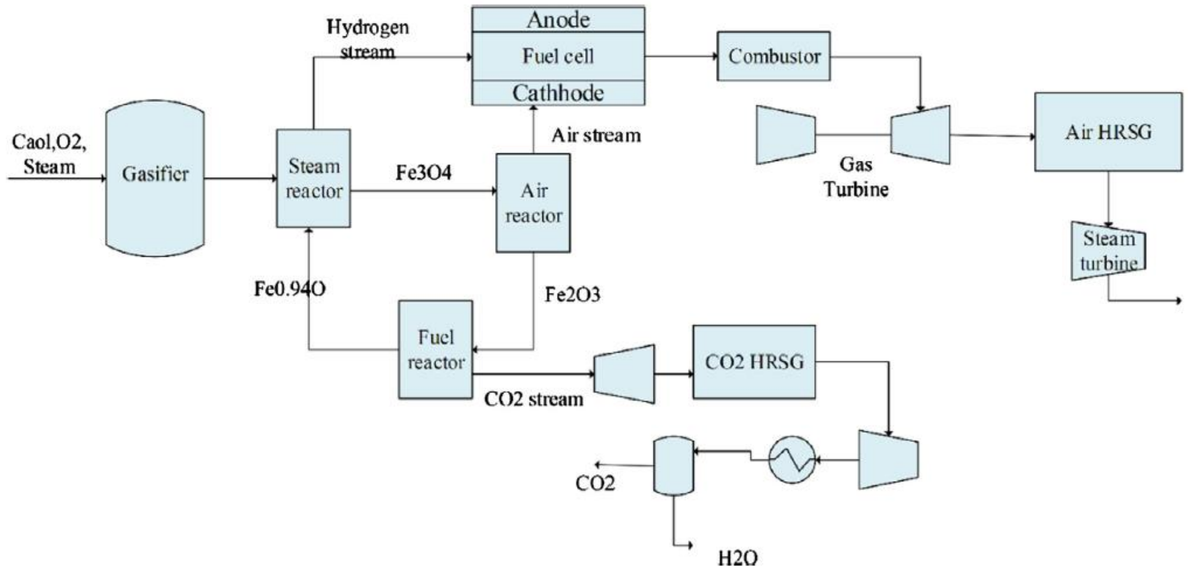


Figure 12: Aghaie's pre- and post-anode chemical looping CO_2 capture process (Aghaie, et al., 2016)

2.4.2. Post-anode CO_2 capture

Post anode CO_2 capture has been extensively studied in SOFC IGFC and natural gas cycles. Common CO_2 capture technologies used are absorption, chemical looping and cryogenic condensation as will be discussed in this section.

A. The Oxy-combustion CO_2 capture process

NETL (NETL, 2009) presented an analysis of the performance and LCOE from two SOFC IGFC power plants. The two plants are differentiated primarily by the operating parameters of the FC Island, one working at approximately atmospheric pressure (1.8 bara), while the other operates at pressurised conditions (18.9 bara) combined with gas-expander power recovery.

These IGFC system configurations include an advanced, catalytic, high-methane gasifier (29% vol. in dry syngas) and syngas cleaning to achieve low sulphur levels in the FC feed gas (0.1 ppm_v). The FC operating parameters were based on previous sub-scale testing (500 mA/cm^2 current density, 0.8V potential and 5 years stack lifetime). The planar FC design provides separation of carbon dioxide (CO_2) as a natural part of operation, while oxy-combustion of the FC effluent enables nearly complete CO_2 capture (99+%).

Overall SOFC operational data and system performance for the two plants as presented in Appendix 1 are:

- Case SO3: Atmospheric-Pressure SOFC IGFC Plant
- Case SO4: Pressurised-SOFC IGFC Plant

Figure 13 shows a simplified flow diagram for the atmospheric IGFC configuration, including the gasifier section, the air separation unit, the gas cleaning section, the FC island, and the steam cycle. For the atmospheric pressure IGFC plant, the net plant power is only 14 MW_e greater than the SOFC power output, so the other generators in the plant (the syngas expander and the steam cycle) essentially counter the plant auxiliary losses. The cathode blower and CO_2 compression train combined account for 66% of the entire parasitic power load (30% and 36%, respectively).

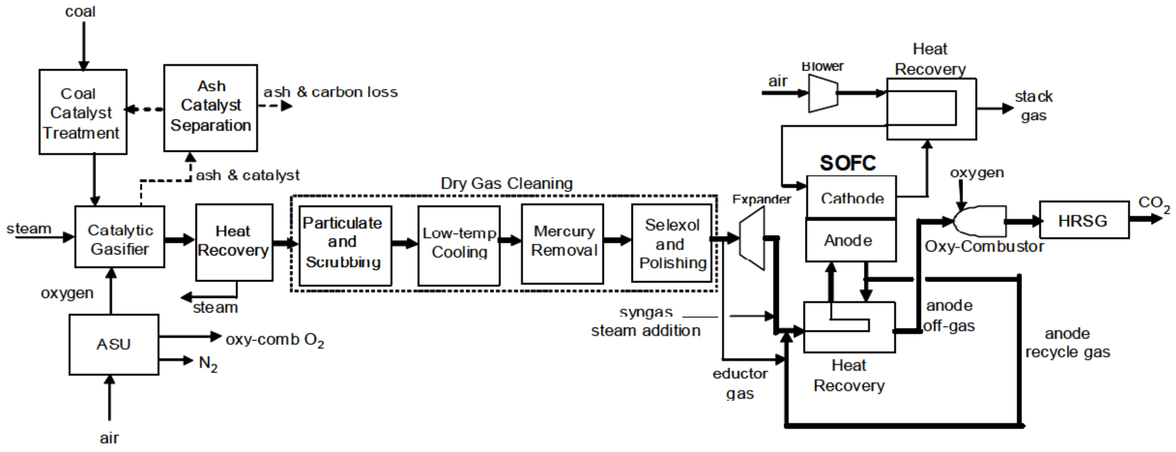


Figure 13: Simplified Flow Diagram of the Atm-Pressure SOFC IGFC Plant (NETL, 2009)

The pressurised IGFC configuration utilises the same type of technologies regarding the coal gasifier section and the air separation section as in the atmospheric pressure IGFC plant. There is no steam bottoming system in the pressurised configuration, with turbine expanders providing power recovery from the SOFC off-gases. A simplified flow diagram is shown in Figure 14. Similar to the atmospheric IGFC, the net plant power is only 32 MW_e greater than the SOFC power output. The cathode and CO_2 compression steps are responsible for approximately 84% of the auxiliary power losses. Even if there is no steam cycle in the pressurised IGFC plant configuration, it has significant power output resulting from the oxy-combustor expander. Overall, the atmospheric SOFC configuration resulted in an efficiency of 49.4%, HHV, net and an LCOE of 88 US$/MWh, while the pressurised configuration presented an efficiency of 56.2%, HHV, net and an LCOE of 79 US$/MWh.

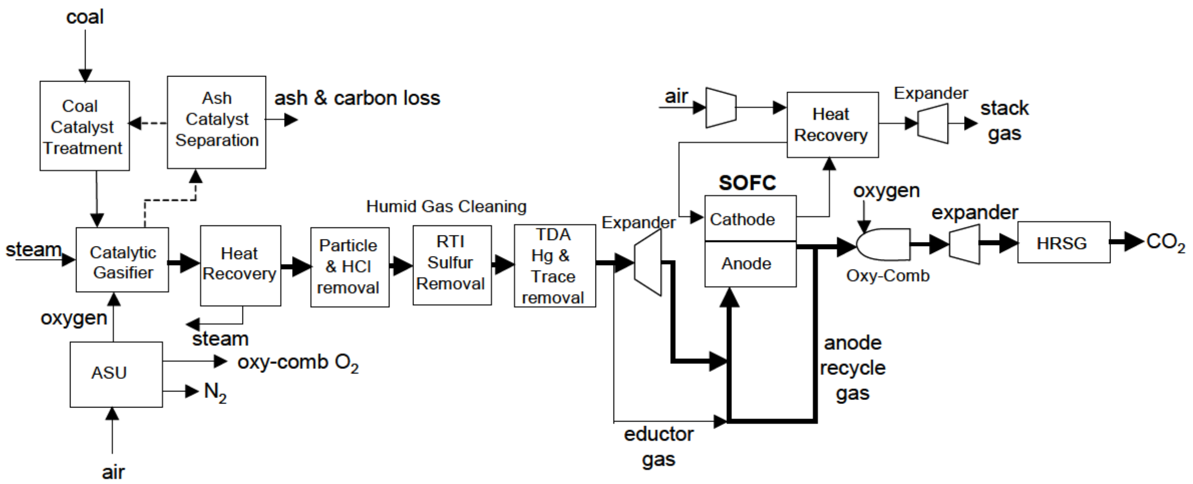


Figure 14: Simplified Flow Diagram of the Pressurized-SOFC IGFC Plant (NETL, 2009)

If those two IGFC plants were operated without the CCS mode, the air separation unit capacity would decrease to less than half, the air separation unit power consumption and the oxygen compression power would be reduced significantly and CO_2 compression power losses would be eliminated. The anode off-gas would be combusted using the cathode off-gas and the resulting product gas exhausted after energy recovery. This would result in efficiencies of up to 53.5% and 61.8% (HHV, net) for the atmospheric and pressurised IGFC cases, respectively.

Grol et al. (Grol, et al., 2008) investigated a pressurised SOFC integrated with a low temperature catalytic gasifier (producing a syngas with a relatively high methane content, ~18%), with/without post-SOFC oxy-combustion CO_2 capture; LHV efficiencies of 59.2%, net and 62.4%, net were reported in the vented and CO_2 capture cases, respectively, however the LCOE was not stated.

Spallina et al. (Spallina, et al., 2011) designed and analysed several IGFC plant configurations, all using a Shell gasifier and a pressurised SOFC, with different gas turbine inlet temperatures (TITs), resulting in LHV efficiencies of 52–54%, net. Spallina et al. extended this work by adding post-SOFC CO_2 capture via oxy-combustion of the anode exhaust, and calculated a 6 percentage point reduction in efficiency. Romano et al. (Romano, et al., 2011) studied a novel IGFC plant configuration that captures CO_2 before the SOFC through physical absorption in a Selexol-based AGR unit; the plant design also features a methanation process upstream of the SOFC (to increase the methane content of the syngas to ~26% vol.), a complete recycle of the shifted anode exhaust to the AGR unit for CO_2 capture, and a state-of-the-art cooled gas turbine capable of exploiting both anode and cathode hot exhaust gases with high efficiency. With this configuration, an efficiency of 51.7%, LHV, net was achieved, 4.5 percentage points higher than that of the plant proposed in Spallina et al.

Duan et al. (Duan, et al., 2013) also investigated various SOFC hybrid systems with CO_2 capture using oxyfuel combustion but focused mainly on methods to maintain the appropriate TIT. Three different methods were investigated, namely steam injection, CO_2 gas injection and heat exchange layout. For a ~142 MW_e SOFC unit the efficiencies reached 62.07%, 63.32% and 63.1%, LHV, net respectively. It was concluded that the SOFC stack, afterburner and HRSG were responsible for the major exergy destruction in the system. Consequently, further research to lower the exergy destruction of the above key units was required in order to enhance the overall system efficiency. Research included optimisation of the operating temperature, the operating pressure and the U_f factor since those are the key parameters that influence the overall system performance.

Lanzini et al. (Lanzini, et al., 2014) analysed the efficiency and economic performance of different configurations of pressurised IGFC plants integrated with a Shell coal gasifier and

CO_2 capture using oxy-combustion. In particular, the three different IGFC configurations studied by Lanzini et al. employed a commercial, pressurised, O_2-blown, entrained flow, dry-feed Shell coal gasifier whose syngas, after processing, feeds an 800°C pressurised SOFC. Post-SOFC CO_2 capture was achieved by combusting the anode exhaust in oxygen to create a hot, pressurised mixture of CO_2 and H_2O, which is expanded and cooled so as to condense out the water. The remaining CO_2 (93%vol.) is compressed and dried for transport and storage purposes. An uncooled gas turbine is used to both compress and heat air to feed the SOFC cathode, and to expand the hot cathode exhaust.

The design variations focus on syngas cleaning and pre-processing upstream of the SOFC power island. In particular, three main system configurations were studied; two with a partial methanation process upstream of the SOFC and one without, as presented in Table A in Appendix A.

- Case SO34: SOFC IGFC Baseline Case (without methanation process)
- Case SO35: SOFC IGFC (with partial methanation process upstream of SOFC)
- Case SO36: SOFC IGFC (with partial methanation process upstream of SOFC)

Depending on the specific plant layout, carbon capture is accomplished either before or after the SOFC power island, or both.

Figure 15 presents the three cases studied by Lanzini et al. The figure is split in three different areas, namely gasifier island, syngas cleaning and processing showing three alternative routes, and power island.

Case SO34 represents the baseline case including a Shell gasifier and a SOFC-based hybrid power cycle. Scrubbed syngas exiting the gasification island enters the AGR unit to remove sulphur compounds. Then the clean syngas is directed to the SOFC power island. This case does not include a syngas methanation step before the SOFC plant.

On the other hand, case SO35 includes a single methanation reactor upstream of the SOFC to increase the methane content of the feed gas. The S-containing syngas exiting the scrubber goes through a partial water-gas shift (WGS) reaction which produces syngas with a H_2/CO molar ratio of 3.0 prior to the methanation step. A fraction of the entering syngas is combined with steam to avoid catalyst coking by achieving a H_2O/CO mole ratio of 2.1, heated to 250°C and then reacted in the first catalytic WGS reactor. The remaining syngas bypasses the reactor and is combined with its effluent and feed water before entering the final stage of a WGS. The shifted syngas is directed to a Rectisol-based absorption process that is able to capture both H_2S and CO_2.

Similarly, case SO36 also employs a methanation unit upstream of the SOFC generator to increase the CH_4 content of the SOFC feed. The catalyst in both cases is Ni-based, being in agreement with previous literature (Udengaard, et al., 2006), but can withstand high CO content syngas as long as a satisfactory amount of steam is fed to the reactor for coking prevention.

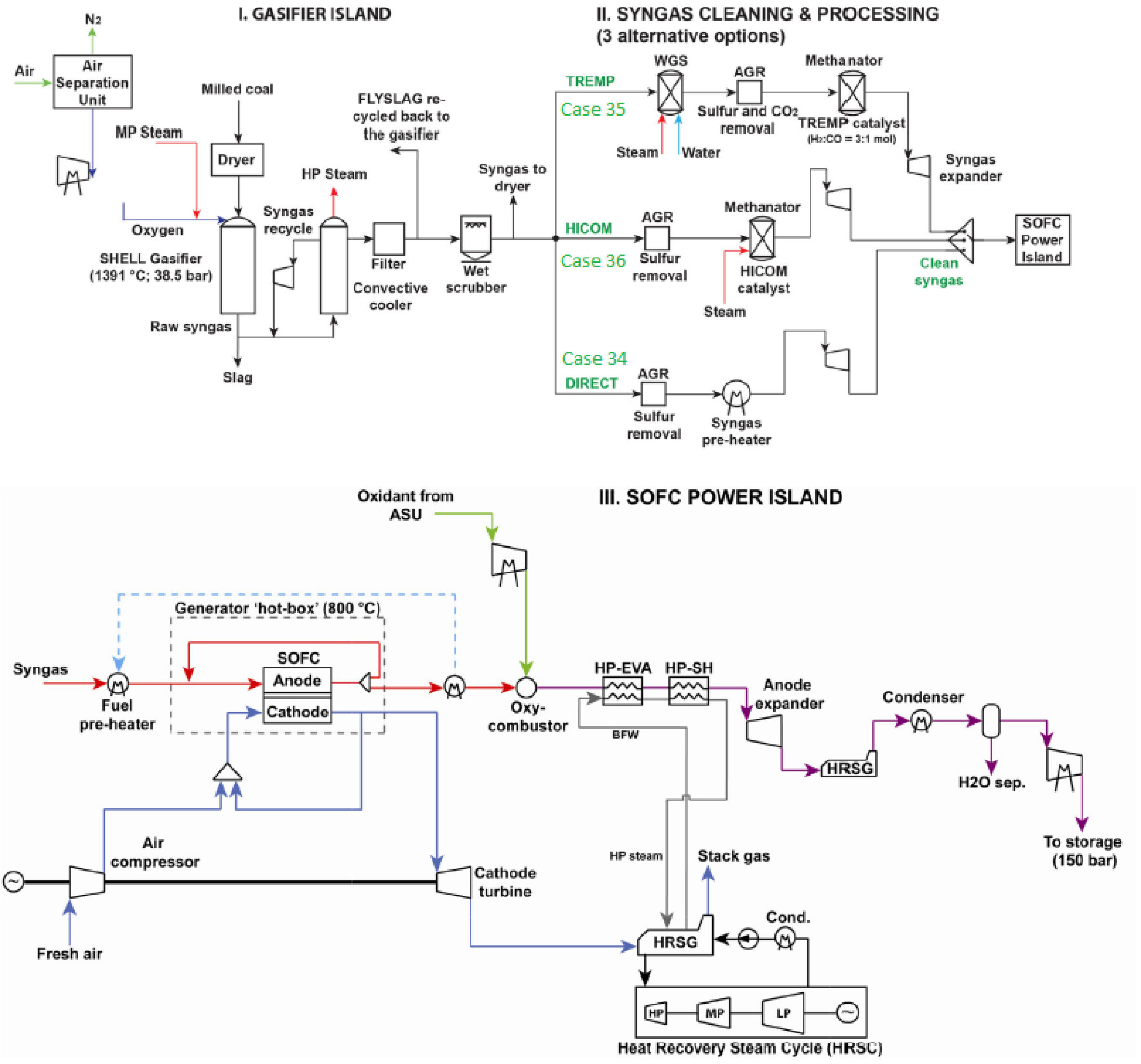


Figure 15: Schematic of the three IGFC plants (Lanzini, et al., 2014)

The SOFC power island is common to all three cases described above, as seen in Figure 15.

The SOFC system assumed in this study is a network of parallel MW-scale modules connected to main pipelines that separately bring air from the GT compressor and fuel from the gasification island, while returning separate cathode and anode exhaust streams to the GT cathode expander and the oxy-combustor, respectively. Each module consists of a pressurised vessel that contains various SOFC stacks, internal recirculators and heat-exchangers for feed stream pre-heating and exhaust flows cooling. The modules are pressurised to approximately 20 bar on both the anode and the cathode sides. Pressurised air for the SOFC cathode inlet is supplied by the air compressor, and the cathode exhaust is expanded in an uncooled turbine. The anode exhaust is oxy-combusted, cooled to 850°C and then expanded to almost atmospheric pressure in an uncooled turbine for additional power generation. Expansion lowers the temperature of the oxy-combusted anode exhaust prior to final cooling and condensation to separate CO_2 from H_2O.

Case SO34 (no methanation) resulted in an overall plant efficiency of 51.4%, HHV, net at an LCOE of 78.8 US$/MWh, whereas case SO36 had a slightly increased efficiency (52.1%, HHV, net) at a lower LCOE (77.6 US$/MWh). In contrast, case SO35 yields both lower efficiency (47.2%, HHV, net) and higher LCOE (89.4 US$/MWh) as it implies the necessity of shifting the syngas and removing CO_2 with an energy-intensive absorption process. Compared to other coal-fired power generating technologies (PC and IGCC), the high capital cost of the SOFC power island in IGFC is offset by its high conversion efficiency and low cost of CO_2 capture process. Looking at the above results it was concluded that at an SOFC module cost of 1700 US$/kW (corresponding to an active cell/stack area cost of 0.054 US$/cm^2), the LCOE of the IGFC case SO36 would be comparable to that of an IGCC-CCS configuration.

Curletti et al. (Curletti, et al., 2015) investigated the techno-economic performance of large integrated biogas SOFC power plants at both atmospheric and pressurised operation with CO_2 capture. Building on a previous project called SOFCOM (Santarelli, et al., 2013) and following (Duan, et al., 2013) findings this study examined different configurations of SOFC plants and a sensitivity analysis regarding U_f, stack temperature and stack pressure was performed. The SOFC module examined in this case produced a constant electrical power of 1MW$_e$ and the reference atmospheric scenario resulted in 48.36%, HHV, net efficiency at US$0.6m NPV. Similarly, the reference case for the pressurised cases resulted in 49.24%, HHV, net and 48.85%, HHV, net efficiency at US$1.63m and US$1.86m NPV for low (2 atm) and high (6 atm) pressure, respectively. The pressurized configurations are able to achieve higher efficiencies than the atmospheric ones essentially because of the production of additional electrical power in gas turbines or expanders.

In a FC, when the U_f increases, a higher concentration of H_2 and O_2 is consumed in the cell resulting in more diluted reactants at the electrode. As a consequence, the reversible voltage decreases because of the decreased concentration of the reactants. This also affects the activation kinetics negatively while it does not influence ohmic and diffusion polarisations: an increase of the ASR is the corresponding effect, being in accordance with the findings reported in the FC Handbook (US DoE, 2004). In contrary, the operating temperature has a positive effect on the ASR because the faster electrochemical and reduced resistivity of the ion conduction electrolyte results in lower stack polarization. The pressurisation of the stack produces an overall increase of the activation kinetics so that higher operating pressures allow the presence of faster electrochemical reactions: this is the reason why, at the same operating conditions, the ASR is lower at pressures higher than atmospheric.

B. The cryogenic process

In the previously considered IGFCs and NGCC integrated with SOFC plant configurations, provided that the CO_2 concentration of the stream at the SOFC anode outlet is approximately 60% on a dry basis, cryogenic CO_2 capture systems can offer a promising option for CO_2 separation from the remaining gases in the mixture (Campanari, et al., 2016). However, regardless of the type of power cycle, the anode exhaust stream exiting the SOFC needs to undergo an additional purification step in order to recover the heating value of the unconverted fuel species and achieve the required CO_2 purity. As a result, apart from the anode off-gas HRSG process, a single high temperature water-gas-shift step is necessary for the gas or steam cycle integration, which converts the residual carbon monoxide to CO_2. The resulting stream is sent to a CO_2 removal section based on a cryogenic process. Residual fuel compounds, mainly consisting of hydrogen, are recycled back to the power plant, while the high purity CO_2 stream is pressurised to supercritical state for transport and storage purposes.

The cryogenic process includes the cooling of the anode stream to a temperature approaching the triple point of CO_2 (56.6°C), where most of the CO_2 condenses and is separated from other fuel species by gravity, since they have a much lower boiling point compared to the CO_2 and therefore remain in the gaseous phase. In general there are two different systems that can be adopted for the cryogenic CO_2 capture process:

- **System A:** a self-refrigerated system where the CO_2 separated in the circuit is the working fluid of the refrigerating cycle required to cool the mixture (Chiesa, et al., 2011);

- **System B:** an external multi evaporation level compression chiller. Since heat has to be removed at temperature below the CO_2 triple point, with a significantly large temperature difference with respect to the ambient temperature, binary refrigerating loops (e.g. propane/ethane or propane/ethylene loops) are generally proposed.

Preliminary techno-economic analysis of the above systems in (Gazzani, et al., 2014) showed that in comparison to **System B** (external chillers), **System A** offered a simpler equipment and plant layout scheme as well as lower operating costs and overall investment.

(Campanari, et al., 2016) studied two plant configurations, with the first utilising an atmospheric pressure SOFC and a Rankine cycle (SOFC + SC) whereas the second was based on a moderately pressurised stack and a Brayton cycle for the heat recovery (SOFC + GT), as seen in Figure 16 and Figure 17.

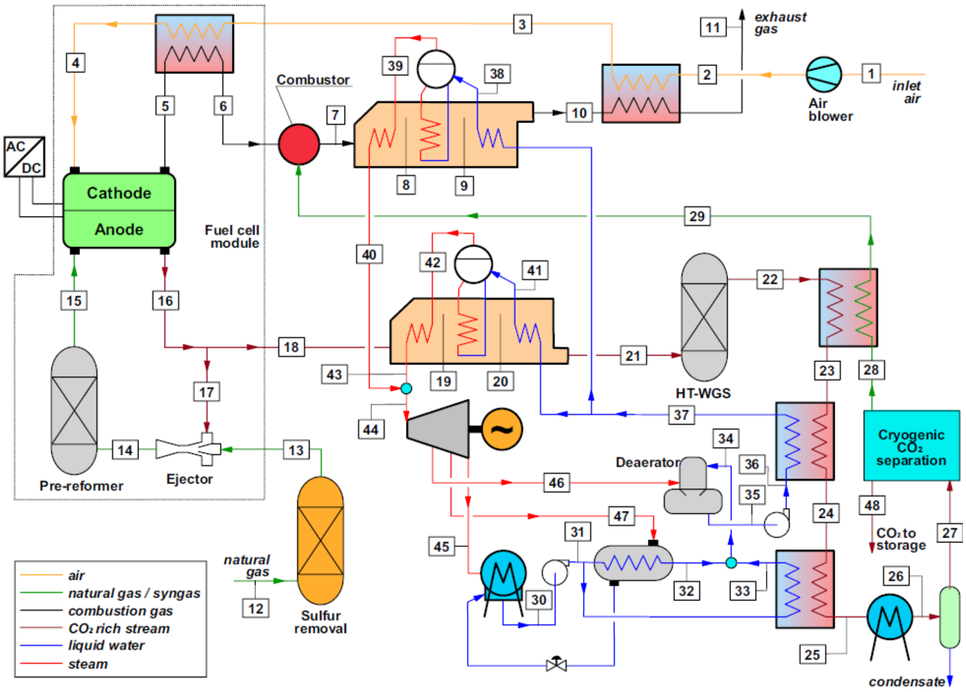


Figure 16: Atmospheric SOFC-based power plant (SOFC + ST) with CO_2 capture section (Campanari, et al., 2016)

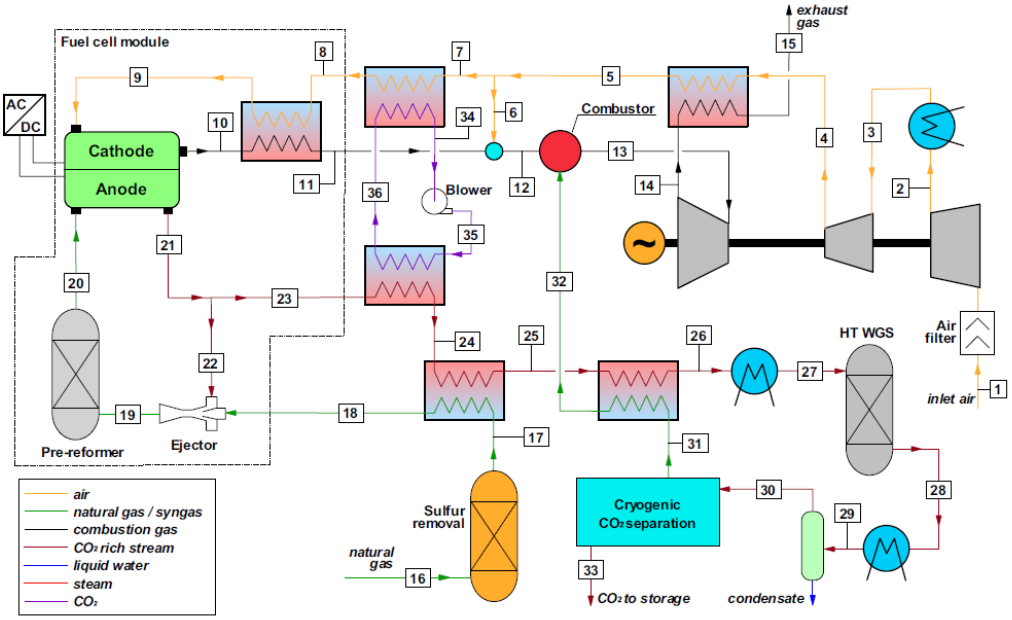


Figure 17: Pressurised SOFC + GT plant with CO_2 capture section (Campanari, et al., 2016)

The achieved net electric efficiency was impressive: 71.56%, LHV for the atmospheric SOFC + SC and 70.82%, LHV for the pressurised SOFC + GT, where the latter could further improve its performance through the addition of an ORC bottoming cycle.

However, a disadvantage of this process is that the CO_2 capture rate is limited to 82.4% for SOFC + SC and 81.4% for SOFC + GT, due to the fact that the cryogenic separation is limited by the phase equilibrium close to the CO_2 triple point. These values are lower than those typically reported from conventional pre or post-combustion CO_2 capture technologies but are very remarkable in terms of plant efficiency. Moreover, a sensitivity analysis showed the possibility of reaching 90% CO_2 capture rates when increasing the U_f up to 92%, having on the other hand a significant efficiency reduction.

2.4.3. Novel post-anode CO_2 capture processes

A novel system configuration was introduced by Samanta and Ghosh (Samanta & Ghosh, 2017). The concept of their study was to repower a thermal power plant incorporating a SOFC and a MCFC upstream and downstream of the present boiler, respectively, as seen in the schematic diagram in Figure 18.

Natural gas is fed to a direct internal reforming (DIR) SOFC module, upstream of the existing boiler plant. The SOFC exhaust is sent to the boiler via burners to supply the equivalent energy of the replaced coal. The SOFC unit consists of a SOFC stack, an afterburner (AB), an inverter, a fuel preheater (FP), an air preheater (APH) and blowers. The residual air and fuel, coming out from cathode side and the anode side, respectively, are combusted in the after-burner. Before directed to the boiler, the output gas of the combustion is used to preheat the fuel and air. The boiler exhaust is sent to the cathode side of an internally reformed MCFC. The MCFC plant consists of a MCFC stack, an air blower, a fuel compressor, heat exchangers (HX), an after-burner (AB) and a HRSG module. Before delivery to the cathode side, air is mixed with flue gas from the boiler and preheated in a HX by the cathode off gas. Natural gas, mixed with steam generated from the HRSG, is sent to the anode of MCFC. The anode off gas is burnt in an after-burner under pure oxygen conditions (98%). An ASU supplies the required oxygen. The outlet of the after-burner is sent through the fuel pre heater (FP), HRSG, moisture separator and finally through the CO_2 compression unit where CO_2 is compressed compression at 110 bar for transport and storage.

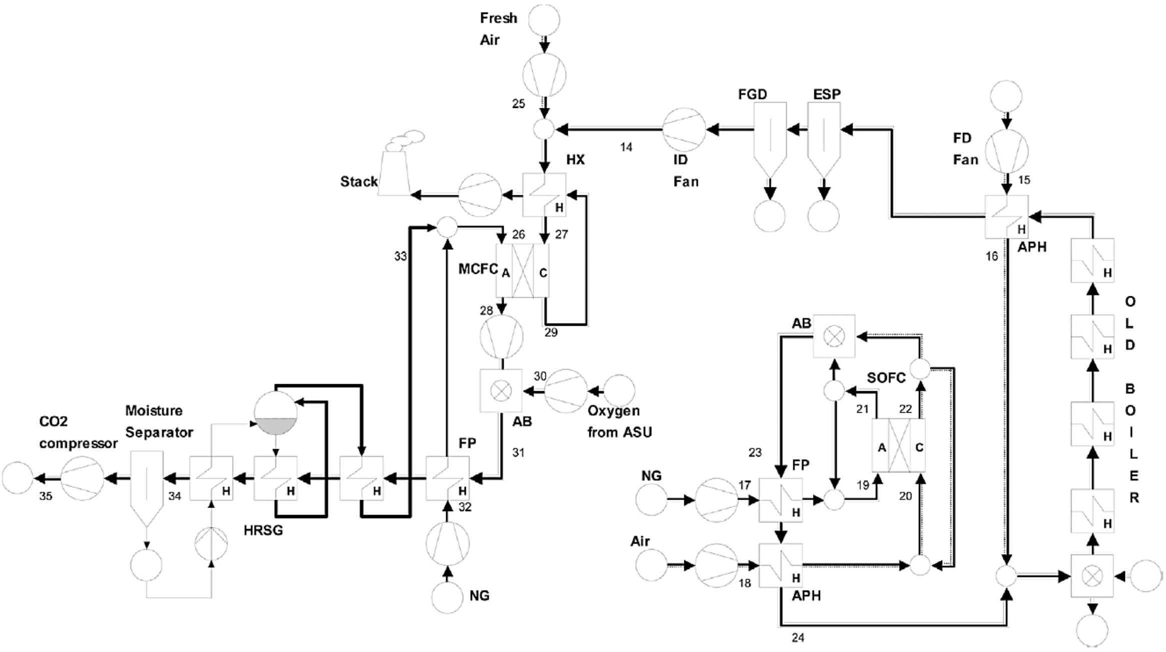


Figure 18: Schematic diagram of repowering a thermal power plant incorporating a SOFC and a MCFC (Samanta & Ghosh, 2017)

Another innovative concept was introduced by (Ziock, et al., n.d.). In this concept an innovative zero-emission coal (ZEC) gasification power generation process was studied, which combined a hydrogasifier, carbonation-calcination reactors and a pressurised SOFC.

Coal is gasified into gas consisting of mainly methane, using hydrogen and water. Methane is then reformed to produce hydrogen through the use of a carbonation reaction, which converts CaO and CO_2 into $CaCO_3$ and supplies energy required to drive the reforming reaction. Hydrogen is partly recycled to the gasifier and the rest is used to produce electricity in a SOFC. The process heat from the SOFC is used to regenerate CaO by calcination of $CaCO_3$, thereby releasing CO_2 that can be stored.

Although very high efficiencies are claimed, both technologies still require large developments to achieve the process integration that is crucial for the technical and economic performance.

The final novel concept that is investigated in this report is SOFC integrated with polygeneration systems and CO_2 capture (Adams & Barton, 2010). Those systems produce both electricity and fuels as major products, creating possibilities of certain synergies that provide efficiency advantages over two separate and independent plants operating in parallel. By adding SOFCs to a polygeneration plant, one can take advantage of the unique properties of SOFCs such as the ability to facilitate 100% CO_2 capture and the fuel flexibility to achieve further benefits.

In particular, Adams and Barton examined 80 different polygeneration systems that produced electricity, methanol, diesel, and gasoline from coal and/or natural gas in various configurations (internal/external reforming) as seen in Figure 19:-

Case A:	Coal only with SOFCs
Case B:	External Natural Gas Reforming with SOFCs
Case C:	Internal Natural Gas Reforming with SOFCs

In Case A of Figure 19, coal is gasified to produce low hydrogen syngas (H_2/CO molar ratio below 1). A portion of the coal-derived syngas is mixed with reformed natural gas (H_2/ CO ratio above 3) to create the desired balance of H_2/CO of about 2. After clean-up and CO_2 removal, the mixed gas is used for methanol synthesis or Fischer−Tropsch synthesis (to produce gasoline and diesel). The rest of the coal-derived syngas is used as fuel to SOFCs (the water-gas-shift step is optional) and after heat recovery, 100% of the resulting CO_2 exiting from the FC is optionally captured.

The external reforming process (Case B) is essentially similar to the coal only process (Case A), with two significant exceptions. First, the WGS reaction is not applied to the coal-derived syngas. In this case the coal-derived syngas undergoes a hydrolysis step in order to convert the COS to H_2S and is then sent to a two-stage Selexol process for H_2S removal. Secondly, after the H_2S clean-up stage and before the removal of CO_2, the 'sweet' coal-derived syngas is mixed with syngas previously produced from a two-stage natural gas reforming process creating a final mixture of a H_2/CO ratio of approximately 2. During the reforming step, natural gas is initially preheated using waste heat and is then directed to a pre-reformer adding also high pressure steam, where 99.9% of the higher hydrocarbons are reformed. Then, after further heating with waste heat, the reactor effluent is reformed in the auto-thermal reformer introducing more steam and O_2 from the ASU, where at least 96% of the methane is converted. The effluent is cooled, the water condensed out, and the resulting dehydrated syngas has a H_2/CO ratio of 3 or higher.

The internal reforming process (Case C follows a similar pattern as case B, however, the natural gas reforming occurs inside the gasifier's radiant cooler, by feeding its tube side with natural gas and water, powered by heat generated from coal gasification. This way, the reforming process cools the coal-derived syngas while producing a valuable high-H_2 syngas from the natural gas. Then, water is condensed out of the natural gas and the dehydrated gas is compressed to 55 bar in the post treatment section before mixing it with coal-derived syngas just before entering the CO_2 removal stage.

Depending on the configuration, using SOFCs for polygeneration improved the total plant efficiency by up to 4 percentage points (for example, 41−45% thermal net system efficiency by HHV) when no CCS was employed, and up to 10 percentage points when CCS was used utilising Selexol (Adams & Barton, 2011). However, the use of SOFCs was only economical depending on the assumptions made for the market prices of fuels (coal and natural gas) and products (electricity and liquids such as methanol, diesel and gasoline). Adams and Barton (2011) compared 80 polygeneration design options using SOFCs to 80 equivalent options using only gas turbines. In general, the use of SOFCs was more economically attractive than purely gas turbine power systems only when the electricity and CO_2 tax rates were generally high and/or the liquid fuel sale prices were generally low.

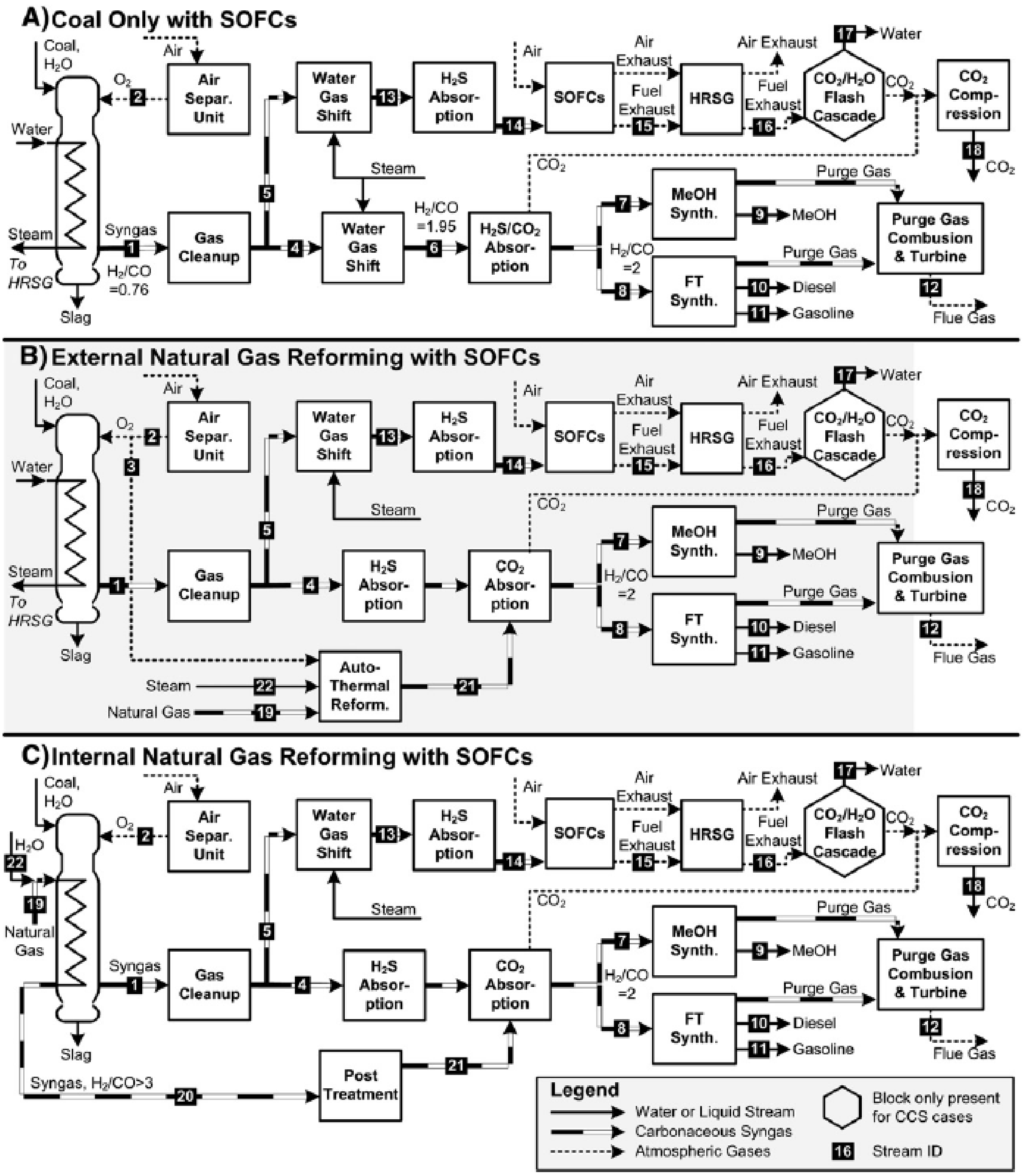


Figure 19: SOFC integrated with polygeneration in three different configurations (Adams & Barton, 2011)

2.5. Options for improving SOFC system performance

A. Higher SOFC working voltage

A higher SOFC working voltage is desirable for two reasons:

a) it makes the SOFC stack more efficient, and

b) a larger portion of the fuel's chemical energy is converted to electric power and less is released as heat.

As a result, an increase of the SOFC operating voltage would help reduce the high operating temperature and consequently reduce the necessary cooling air flow rate and the associated compression power.

With the progress of SOFC manufacturing and the improved performance reported in recent studies, a higher working voltage can be expected to be used in the future, compared to the operating voltage reported in this report. As seen in Table A of Appendix 1, most studies used a voltage between 0.7 and 0.8V, however, if it is possible to achieve values greater than 0.8V then there could be significant benefits.

A study by Li et al. (Li, et al., 2011) proved that when the cell working voltage is increased from 0.8V to 0.85V, the system thermal efficiency improves from 42.1% to 47.8%, HHV, net due to the relatively higher power output of the SOFC stack and due to the relatively smaller parasitic power consumption of the SOFC air compressor.

One drawback of the increase of the SOFC working voltage is that the anode exhaust temperature becomes lower. This makes it more challenging to pre-heat the fuel stream with the anode exhaust. Another parameter that needs to be taken into consideration is that by increasing the SOFC working voltage the average current density decreases which would mean that for a constant total power output more SOFC stacks will be needed, resulting in higher capital cost. As a result, the optimum cell working voltage does not only depend on performance parameters but also on economic factors. It should also be noted that due to the theoretical limit of the Nernst potential, there is a certain constraint of further increasing the SOFC working voltage.

B. Cascading of identical SOFC stacks: a viable design strategy

As discussed above, a key factor for achieving high thermal efficiency is to reduce the overall cooling air flow rate (increase air utilisation) while satisfying the SOFC stack safe operating temperature requirements. An innovative SOFC module design strategy was suggested by (Li, et al., 2011), which comprises of cascading identical SOFC stacks with series air flow, parallel fuel flow, and intra-stack introduction of fresh air to produce roughly identical operating conditions for each stack in the module. The proposed design may be able to achieve high IGFC system efficiencies as

a) air utilization in each stack can be held low enough to effectively control the internal SOFC temperature peak;

b) the cathode exhaust coming out of one SOFC stack still contains large amount of O_2, which can be used in downstream stacks; and

c) typically for the counter-flow configuration the cathode exhaust temperature is not significantly higher than the air inlet temperature, which assures that the cathode exhaust coming from upstream SOFC can still effectively work as a cooling flow in downstream SOFC stacks.

In addition, fresh make-up air can be added in after each stack to provide more O_2, to cool the air flow to the desired temperature, and to provide roughly identical operating conditions for each stack. This design feature of the developed SOFC stack module concept is important to the cost of the system, which will depend upon the mass manufacturing of identical stack modules.

An example of such a design is shown in the Sankey diagram of Figure 20. This design comprises of four stages of cascading SOFC stacks with similar design parameters (geometries, cell properties, etc.) and a working voltage of 0.8V for each stack. The width of the streams in the Sankey diagram represents the mass flow rate of each stream. The total incoming fuel flow is pre-heated to 650°C and is evenly distributed to the four SOFC stacks. The air flow is cascaded in series through the four stages of SOFC stacks, with the main air flow at the first stack being also preheated to 650°C. Small amounts of cold fresh make-up air are mixed with the cathode exhausts from stacks I, II and III, to cool down the main air flows lightly as well as to provide more O_2 for electrochemical reaction in downstream stages. There is no need to pre-heat the make-up air because relatively lower temperature is desired for cooling purposes; thus make-up air coming out of the air compressor at about 330°C can be used directly.

By employing this design, the overall air utilisation factor of the SOFC stacks can be increased significantly. The parasitic power consumption of the SOFC air compressor reduces significantly, due to the much smaller overall cooling air flow rate. In addition, the last stage cathode exhaust temperature and the anode exhaust mixture temperature are significantly higher than those achieved in the conventional SOFC designs, which is beneficial to the IGFC system thermal optimisation and operation of the pre-heaters.

The system manages to achieve a thermal efficiency of 55.2%, HHV, net which is higher than the systems with the increased operating voltage described above.

A drawback of the proposed design is that some pressure head of the compressed flows is wasted because of the cascading stages. However, the performance of the IGFC with cascading SOFC stacks is considered satisfactory and more research is needed to validate the above reported results in terms of performance, in addition there are no details on the economic performance which would allow assessment against other suggested performance improvements.

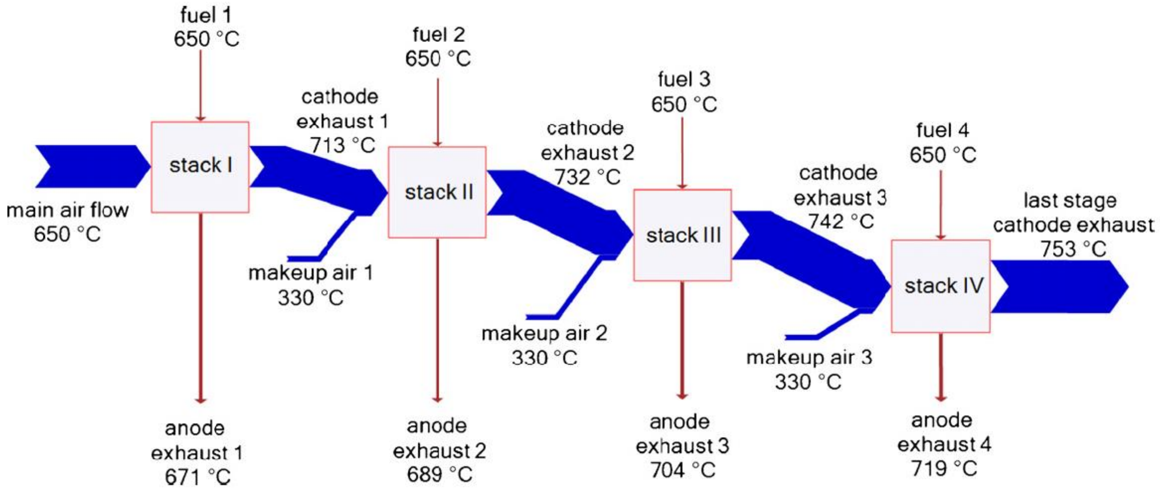


Figure 20: Sankey diagram of the SOFC module design with four stages of cascading SOFC stacks (Li, et al., 2011)

3. MOLTEN CARBONATE FUEL CELLS

3.1. Background of MCFC technology

The MCFC system consists of individual cells which are stacked to make up any desired power, with scales ranging from small kW_e to large MW_e applications (US DoE, 2004).

The individual cell consists of an anode and a cathode where the conversion reactions take place, joined by an electrolyte which closes the electrical circuit, as seen in Figure 21.

Both anode and cathode are nickel-based whereas the electrolyte consists of harmless salts of lithium, potassium and sodium carbonates in molten state and are suspended in a porous ceramic matrix.

The high operating temperature (650°C) and nickel anode are also excellent conditions for reforming and shift reactions which firstly convert hydrocarbon species and water into H_2, then further converts water and carbon monoxide into additional hydrogen and CO_2. As a consequence, the MCFC can operate with both pure hydrogen as well as hydrocarbons, with the oxygen for the final conversion of hydrogen being transported across the electrolyte in the form of CO_3 ions, generating electrical current.

It is interesting to note that CO_2 is necessary for the operation of the MCFCs as a closed-loop reagent: the CO_2 is consumed at the cathode (together with oxygen) at the same rate at which it is released at the anode. This role of CO_2 makes MCFC systems ideal for CO_2 separation applications from the flue gas of combustion-based power plants.

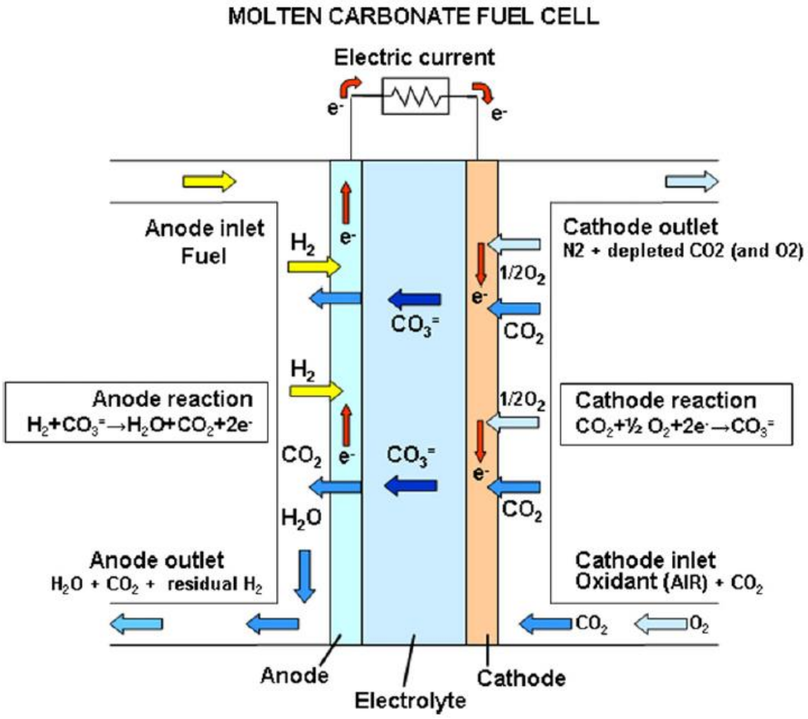


Figure 21: The Molten Carbonate FC (Caprile, et al., 2011)

As mentioned previously, the high temperature MCFC can be fed with any hydrogen-carbon mixture as fuel, therefore a variety of fuels such as natural gas, biogas, syngas from gasified coal, waste or biomass, but even liquid fuels such as ethanol can be adopted. The high operating temperature of the MCFC helps to process all these different fuels, but for safe and enduring operation of the MCFC a careful clean-up of the fuel is necessary beforehand, since FC systems are less tolerant to impurities than combustion based systems.

Figure 21 presents the operating configuration of the MCFC. Excluding any prior reforming and shift reactions which also take place in the cell, the half-cell electrochemical reactions are:-

$$H_2 + CO_3^{2-} \rightarrow H_2O + CO_2 + 2e^- \tag{3-1}$$

at the anode, while at the cathode the reaction is:

$$\tfrac{1}{2}O_2 + CO_2 + 2e^- \rightarrow CO_3^{2-} \tag{3-2}$$

The overall cell reaction[1] is:

$$H_2 + \tfrac{1}{2}O_2 + CO_2 \text{ (cathode)} \rightarrow H_2O + CO_2 \text{ (anode)} \tag{3-3}$$

Apart from the reaction between H_2 and O_2 to H_2O, Equation 2-3 shows a transfer of CO_2 from the cathode gas stream to the anode gas stream via the formation of the CO_3^{2-} ion through the electrolyte. The reversible potential of an MCFC, taking into consideration CO_2 transfer, is given by Equation (3-4):-

$$E = E^0 + \frac{RT}{2F} ln \frac{P_{H_2} P_{O_2}^{1/2}}{P_{H_2O}} + \frac{RT}{2F} ln \frac{P_{CO_{2,c}}}{P_{CO_{2,a}}} \tag{3-4}$$

where the subscripts a and c refer to the anode and cathode gas compartments, respectively. When the partial pressures of CO_2 are identical at the anode and cathode, and the electrolyte is invariant, the cell potential depends only on the partial pressures of H_2, O_2, and H_2O. Typically, the CO_2 partial pressures are different in the two electrode compartments and the cell potential is affected accordingly.

Factors affecting the selection of operating conditions of the MCFC are stack size, heat transfer rate, voltage level, load requirement and cost. The performance curve is defined by cell pressure, temperature, gas composition and U_f. Typical MCFCs will generally operate in the range of 100 to 200 mA/cm^2 at 750 to 900 mV/cell.

A. Effect of Pressure

The dependence of reversible cell potential on pressure is evident from the Nernst equation. For a change in pressure from P_1 to P_2, the change in reversible potential (ΔV_p) is given by

$$\Delta V_P = \frac{RT}{2F} ln \frac{P_{1,a}}{P_{2,a}} + \frac{RT}{2F} ln \frac{P_{2,c}^{3/2}}{P_{1,c}^{3/2}} \tag{3-5}$$

where the subscripts a and c refer to the anode and cathode, respectively. In an MCFC with the anode and cathode compartments at the same pressure (i.e., $P_1 = P_{1,a} = P_{1,c}$ and $P_2 = P_{2,a} = P_{2,c}$):-

$$\Delta V_P = \frac{RT}{2F} ln \frac{P_1}{P_2} + \frac{RT}{2F} ln \frac{P_2^{3/2}}{P_1^{3/2}} = \frac{RT}{4F} ln \frac{P_2}{P_1} \tag{3-6}$$

At 650°C the above equation can be simplified to:-

$$\Delta V_P(mV) = 20 ln \frac{P_2}{P_1} = 46 \log \frac{P_2}{P_1} \tag{3-7}$$

---

[1] CO is not directly used by electrochemical oxidation, but produces additional H_2 when combined with water in the water gas shift reaction.

which means that a tenfold increase in cell pressure corresponds to an increase of 46 mV in the reversible cell potential at 650°C. Increasing the operating pressure of the MCFC results in enhanced cell voltages because of the increase in the partial pressure of the reactants, gas solubility's and mass transport rates. However, some undesirable effects that also stem from increasing the operating pressure of the cell are side reactions such as carbon deposition (Boudouard reaction):-

$2CO \rightarrow C + CO_2$ (3-8)

and methane formation (methanation):-

$CO + 3H_2 \rightarrow CH_4 + H_2O$ (3-9)

In addition, decomposition of CH_4 to carbon and H_2 is possible following the reaction below:-

$CH_4 \rightarrow C + 2H_2$ (3-10)

but this reaction is suppressed at higher pressure. According to Le Chatelier's principle, an increase in pressure will favour carbon deposition by Equation (3-8) and methane formation by Equations 3-9 and 3-10. The water-gas shift reaction is not affected by an increase in pressure because the number of moles of gaseous reactants and products in the reaction is identical. Ideally, carbon deposition in a MCFC needs to be avoided as it can lead to plugging of the gas passages in the anode and subsequently to stack degradation. Methane formation is also detrimental to cell performance because the formation of each mole consumes three moles of H_2, which represents a considerable loss of reactant which inevitably would reduce power plant efficiency.

B. Effect of Reactant Gas Composition and Utilisation

The composition of the reactant gases influences directly the voltage of the MCFCs due to the presence of CO at the water gas shift reaction at the anode and the consumption of both CO_2 and O_2 at the cathode. Previous studies have shown that an increase in the reactant gas utilisation results in a decrease in the cell performance (Farooque, 1991). As the reactant gases pass through the cell, the cell voltage decreases as a result of polarisation (i.e. activation, concentration) and changing gas composition and partial pressures.

Regarding the cathode side of the MCFC, as seen from equation 3-2, the electrochemical reaction involves the consumption of two moles CO_2 for each mole of O_2. This ratio provides the optimum cathode performance, however, as this ratio decreases the cathode performance also decreases and a limiting current is apparent. For the anode side the anode potential varies as a function of the $[H_2]/[H_2O][CO_2]$ ratio and a higher potential is obtained when this ratio is higher, as discussed in (US DoE, 2004).

C. Effect of Impurities

If the main source of fuel for the MCFC is expected to be syngas derived mainly from gasified coal it expected to contain various contaminants that need to be pre-treated before reaching the anode side of the MCFC. Various studies have been performed to evaluate the concentration levels that can be tolerated by MCFCs without significant degradation in performance or reduction in cell life. A list of possible effects of contaminants from coal-derived fuel gases on MCFCs is summarized in Table 3 (Anderson & Garrigan, 1984).

Table 3: List of contaminants and their effects on the MCFC stack (US DoE, 2004)

Class	**Contaminant**	**Potential Effect**
Particulates	Coal fines, ashes	Plugging of gas passages
Sulphur compounds	H_2S, COS, CS_2, C_4H_4S	- Voltage losses - Reaction with electrolyte via SO_2
Halides	HCl, HF, HBr, $SnCl_2$	- Corrosion - Reaction with electrolyte
Nitrogen compounds	NH_3, HCN, N_2	- Reaction via electrolyte via NOx
Trace metals	As, Pb, Hg, Cd, Sn Zn, H_2Se, H_2Te, AsH_3	- Deposits on the electrode - Reaction with the electrolyte
Hydrocarbons	C_6H_6, $C_{10}H_8$, $C_{14}H_{10}$	- Carbon deposition

D. Voltage Output and Cell Life

The aim of a FC should be to maximise thermal or exergetic efficiency while minimising cell stack degradation in order to produce as much power for as much time as possible. These three targets can be achieved by improving the FC design (optimisation of electrolytes and electrocatalysts, improvements in electrode structures, thinner cell components, etc.) and/or by adjusting the operating conditions (e.g., higher temperature, higher gas pressure, and change in gas composition to lower the contaminant concentration), as discussed in (Farooque, 1990).

The voltage output of a MCFC is reduced by ohmic, activation, and concentration losses that increase with increasing current density. As shown in Figure 22, the activation polarisation (reaction rate loss) is more important at lower current densities. At this point, electronic barriers must be overcome prior to ion and current flow. Resistance loss (Ohmic polarisation) is directly affected by current, increasing over the entire range of current since cell resistance remains essentially constant. Concentration polarisation (gas transport loss) takes place over the entire range of current density, but they become significant at high limiting currents where it is more difficult to provide enough reactant flow to the cell reaction sites.

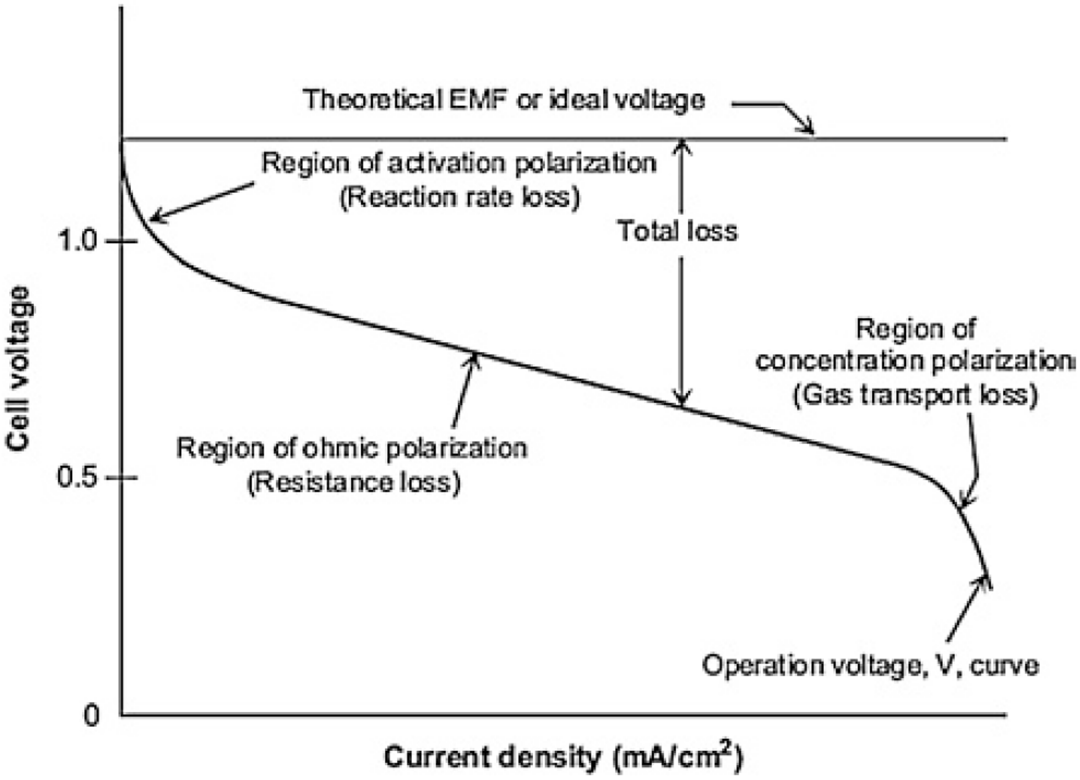


Figure 22: Current density vs cell voltage (Farooque, 1990)

Changing the cell operating parameters (temperature and pressure) can present both advantages and disadvantages regarding the impact on the performance of the FC and compromises in the operating parameters are essential to meet the application requirements of lower system cost and acceptable cell life.

Regarding MCFC's performance in practical operating conditions, the following potential balance must be approximately satisfied in the cell:-

$$V = E - n_{ne} - (R_{ir} + R_a + R_c)J \qquad (3\text{-}11)$$

Here V is the output voltage, E the equilibrium potential (i.e. open circuit voltage), n_{ne} Nernst loss, R_{ir} the internal resistance, R_a and R_c the anode and cathode reaction resistance due to the polarization in both electrodes and J the current density. The decay of the output voltage according to time t is expressed in the following equation:-

$$V(t) = E - n_{ne} - \left(R_{ir}(t) + R_{a(t)} + R_c(t)\right)J \qquad (3\text{-}12)$$

The variation of the output voltage $V(t)$ can be represented by the increase of R_{ir}, R_a and R_c over time.

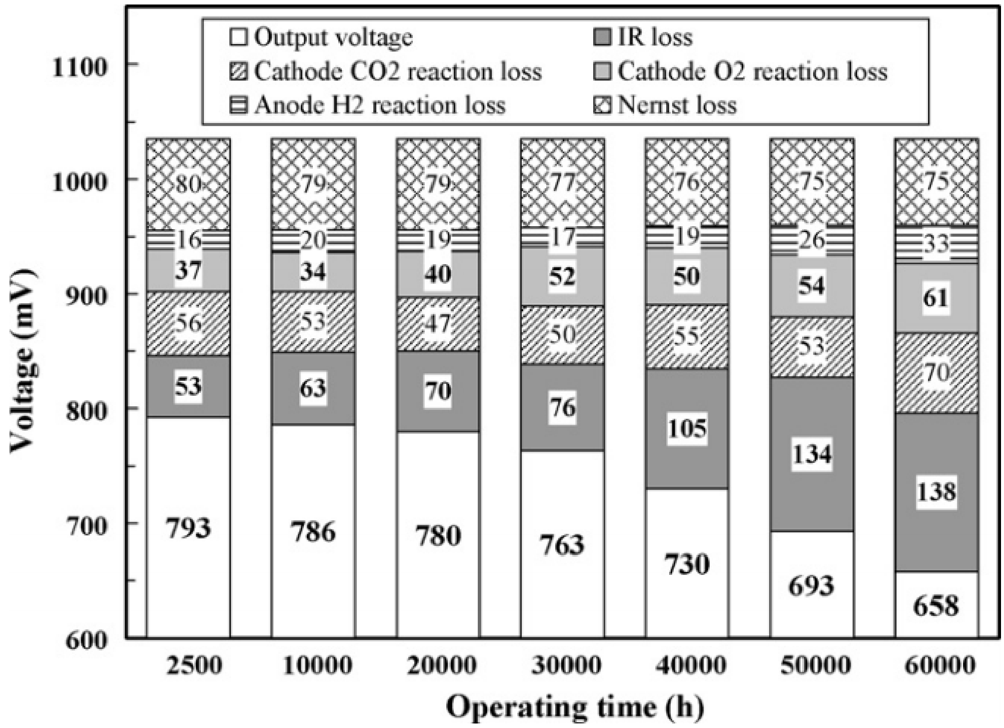


Figure 23: Analysis of a cell performance at 600°C versus operating time (Morita, et al., 2010)

Figure 23 shows an analysis of a MCFC performance at 600°C versus operating time studied by (Morita, et al., 2010). The results suggest that the main factor responsible for the voltage degradation is the increase of internal resistance R_{ir}. The second factor is the cathode CO_2 and O_2 reaction losses, however it is evident that the cathode O_2 reaction losses increase significantly over time, especially after 30,000h. The reason for this could be the lower solubility of oxygen in the carbonate melt in comparison to hydrogen or carbon dioxide. Regarding the anode H_2 reaction loss, an increase was observed after 50,000h, which can be explained by the increase of the gas cross leakage. The aging cell components, such as electrodes and electrolyte matrix, change the structure of the gas–liquid contact inside the porous electrode where the electrochemical reaction occurs. The change of the structure would be advantageous to increase the cathode oxygen reaction since the amount of reactive oxygen dissolved in the melt is smaller than that of the hydrogen or carbon dioxide.

Stack tolerance is a critical matter for the commercialisation of MCFCs. Sufficient cell performance must be maintained over the desired length of service; MCFC manufacturers claim that an acceptable average potential degradation should be no greater than 2mV per 1,000 hours of operation over a cell stack lifetime of 40,000 hours (US DoE, 2004).

E. Internal Reforming

As discussed previously for SOFC, the concept of internal reforming is highly practical for systems that operate at high temperatures (SOFC / MCFC) where the steam reforming reaction can be sustained with catalysts. The internal reforming FC eliminates the need for external fuel processors, minimising capital costs and system complexity. In earlier studies it was proven that the internal reforming MCFC configuration can provide a highly efficient, simple, reliable, and cost effective alternative to the syngas-fed MCFC system (US DoE, 2004).

There are two alternate approaches to internal reforming of carbonaceous fuels, namely direct internal reforming (DIR) and indirect internal reforming (IIR).

For the case of IIR, the reformer section is separate, but adjacent to the FC anode. The cell takes advantage of the close-coupled thermal benefit where the exothermic heat of the cell reaction can be used for the endothermic reforming reaction. Another advantage is that the reformer and the cell environments do not have a direct physical effect on each other. A disadvantage is that the conversion of methane to hydrogen is not promoted as well as in the direct approach.

Regarding the DIR case, hydrogen consumption reduces its partial pressure, thus driving the methane reforming reaction to the right.

3.2. MCFC for CO_2 capture

As discussed in Section 3.1, MCFCs utilise carbonate ions as the charge carriers for the cathode side. Their operating temperature varies between 550°C and 650°C in order to maintain a sufficient electrolyte conductivity and chemical reactivity without the use of noble metal catalysts. MCFC can simultaneously act as a 'CO_2 concentrators' and a 'power producer' (Campanari, et al., 2010). Utilisation of a MCFC as a carbon dioxide concentrator was investigated experimentally by Sugiura et al. (Sugiura, et al., 2003), where these proved that the experimental values for the CO_2 removal rate from cathode to anode performed in a MCFC almost matched theoretical calculations.

A key conclusion from their work is that the ratio between the partial pressures of oxygen and carbon dioxide of the cathode directly affects the performance of the cell, being in accordance with the relevant discussion in the previous section. Later, Campanari et al. (Campanari, et al., 2010) studied the separation of CO_2 from NGCCs integrated with MCFC operating at the back-end of their exhaust stream. Results showed that a significant reduction of CO_2 of up to 80% with negligible impact on electrical efficiency is possible. This reduction in CO_2 emissions was achieved with a MCFC sized to contribute only 20% of the net power generation. A similar research studied the possibility of utilising cryogenic separation of CO_2 from combined cycles integrated with MCFC (Chiesa, et al., 2011). The possibility of recirculating a fraction of the CO_2 recovered in the cathode was investigated, thereby controlling the utilisation factor of CO_2. This way separation via oxy-fuel combustion presented in Campanari et al. can be avoided.

Another publication from the same research group (Sanchez, et al., 2013) compared three configurations for integrating a MCFC with an internal combustion engine. Carbon emissions from the combined system were reduced significantly, however, a separate research came to the conclusion that these types of plants are un-profitable at the moment. However, it was also stated that future cost reductions and incentives for FCs and avoided carbon emissions could make this an attractive technology. The above economic analysis considered a small scale MCFC plant integrated with a CHP system able to separate 63% of the inlet carbon quantity at an overall electrical efficiency of 35%, HHV, net. Further improvements in performance can improve the financial feasibility of the MCFC - CO_2 capture concept as described in the following cases.

Following the above initial studies, Rexed et al. (Rexed, et al., 2015) focused on a lab-scale MCFC system fed with simulated flue gas, in order to show the feasibility of MCFC as a CO_2 separator for gas turbine and coal-fired power plants. Milewski et al. (Milewski, et al., 2009) showed that the use of a MCFC can reduce CO_2 emissions from a PC power plant by 61%. Manzolini et al. (Manzolini, et al., 2012) showed that integration of MCFCs in NGCCs increases plant output by 40% while reducing CO_2 emissions by 70–80%. Mamaghani et al. (Mamaghani, et al., 2015) published an energetic, exergetic, economic and environmental analysis of a hybrid MCFC-GT configuration in an attempt to optimise the proposed system. Lastly, Campanari et al. (Campanari, et al., 2014) published a techno-economic assessment of capturing CO_2 from NGCC power plants using MCFC technology and its effect on cost of electricity and CO_2 capture cost. Some of the most important findings will be presented in the next sections.

3.2.1. NGCC integrated with MCFC plants

This section focuses on NGCC power cycles with a MCFC placed downstream of the gas turbine and upstream of the heat recovery steam generator (HRSG). The exhaust gases from the gas turbine are used as the feed for the cathode of the MCFC, where CO_2 is transferred to the anode side, concentrating the CO_2 in the anode effluent. As already mentioned, internal reforming is a preferable approach for the MCFC; natural gas feeding the MCFC must be desulphurised since reformer catalysts and MCFCs do not tolerate the presence of sulphur compounds above 0.5–1 ppm_v. Various approaches, including zinc-oxide absorption beds or activated carbon filters, are effective for natural gas desulphurisation.

After the FC, the anode effluent requires additional purification to recover the unconverted fuel species and achieve the requested CO_2 purity (i.e. >96%). Below, two different NGCC–MCFC plant configurations are compared, based on two different CO_2 separation processes, namely cryogenic and oxy-combustion processes (Del Piano, 2012):-

A. Cryogenic option

The first approach for using MCFC for CO_2 capture in a NGCC power plant is the cryogenic option, provided that the CO_2 concentration at MCFC anode outlet is around 80% on a dry basis, separating CO_2 from the non-condensable species included in the anode exhaust mixture.

In this option the anode stream is cooled down to a temperature approaching the triple point of CO_2 (-56.6°C), until most of the CO_2 condenses and separates gravitationally. At that temperature other fuel species with much lower boiling point remain in the gas phase. This process could reach an 89.3% CO_2 separation efficiency, delivering a CO_2 stream of 98.8% mol purity. In addition to CO_2 separation, the remaining exhaust stream, that contains some combustible species, is sent to the gas turbine, along with natural gas. This natural gas input represents about 9% of the total fuel input (LHV basis), increasing slightly the CO_2 fraction at GT outlet (5% vs. 4% without CO_2 capture).

This option is shown in Figure 24, together with some examples of thermodynamic conditions and chemical composition in relevant points. For a complete list of properties the reader could refer to (Campanari, et al., 2014). This configuration resulted in a 58.88%, HHV, net efficiency and the reported LCOE was 84.66 €/MWh at 120.4 €/Tonne of CO_2 avoided – refer to Case MC1 of Table B in Appendix 1.

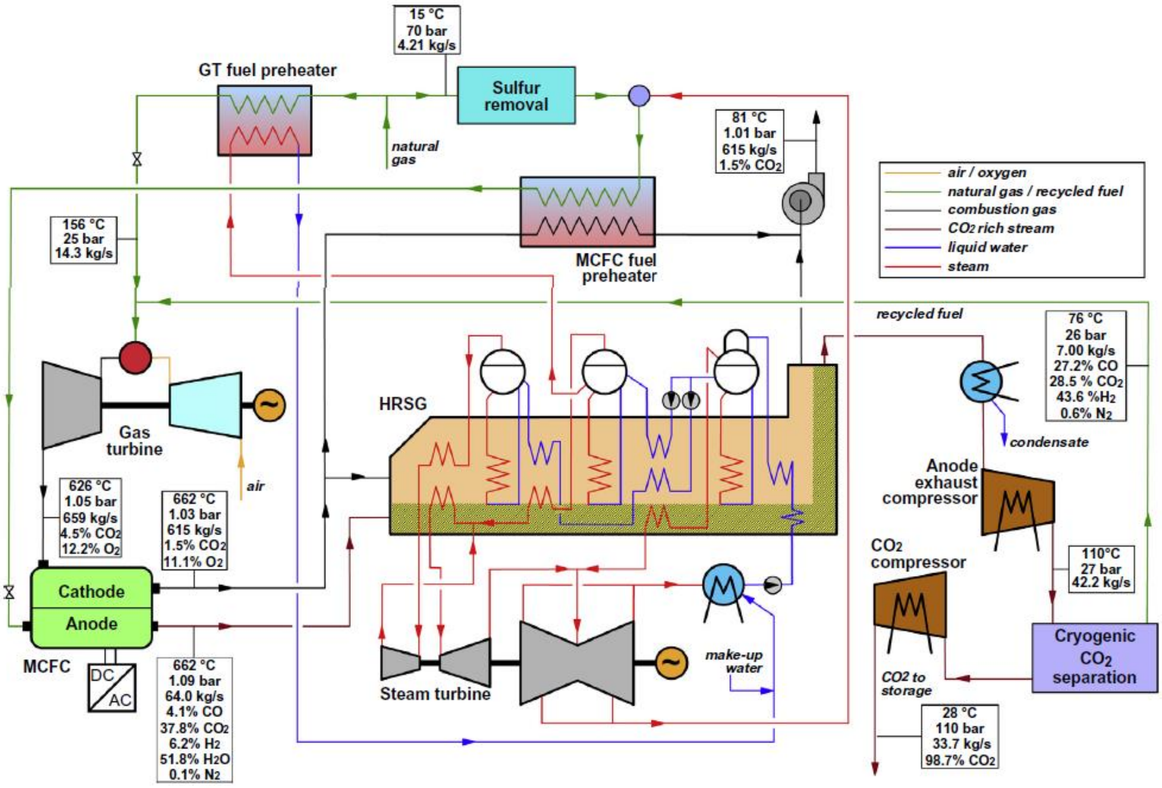


Figure 24: NGCC + MCFC plant configuration with cryogenic CO_2 capture (Cryo) (Campanari, et al., 2014)

B. Oxy-combustion option (Oxy)

In the second approach, residual fuel components are combusted in a boiler with near-pure oxygen, stemming from an air separation unit (ASU), to avoid CO_2 dilution. It is assumed that the ASU produces oxygen of 98% purity (the rest being 0.67% N_2 and 1.33% Ar) at atmospheric pressure through a cryogenic, double-column air separation process. The resulting combustion products are composed mainly of H_2O and CO_2 and they are cooled down to recover heat for steam generation; after water condensation, the fraction of incondensable species in dry CO_2 is below 4%.

This option is shown in Figure 25. This configuration resulted in a 57.53% (HHV) net efficiency and the reported LCOE was 81.53 €/MWh at 107.7 €/Tonne of CO_2 avoided – see case MC2 of table B in Appendix 1.

In both cases the high purity CO_2 stream recovered by the above separation processes is compressed above supercritical pressure for subsequent transport and storage. Other possible configurations could include: an external reforming process for the MCFC (Manzolini, et al., 2012), or partial recycle of the stream exiting the HRSG to the inlet of gas turbine compressor in order to increase the CO_2 concentration in the exhaust gases. These solutions have been investigated in previous works (Campanari, et al., 2013), each showing advantages and disadvantages. However, they generally present higher complexity and will not be discussed here.

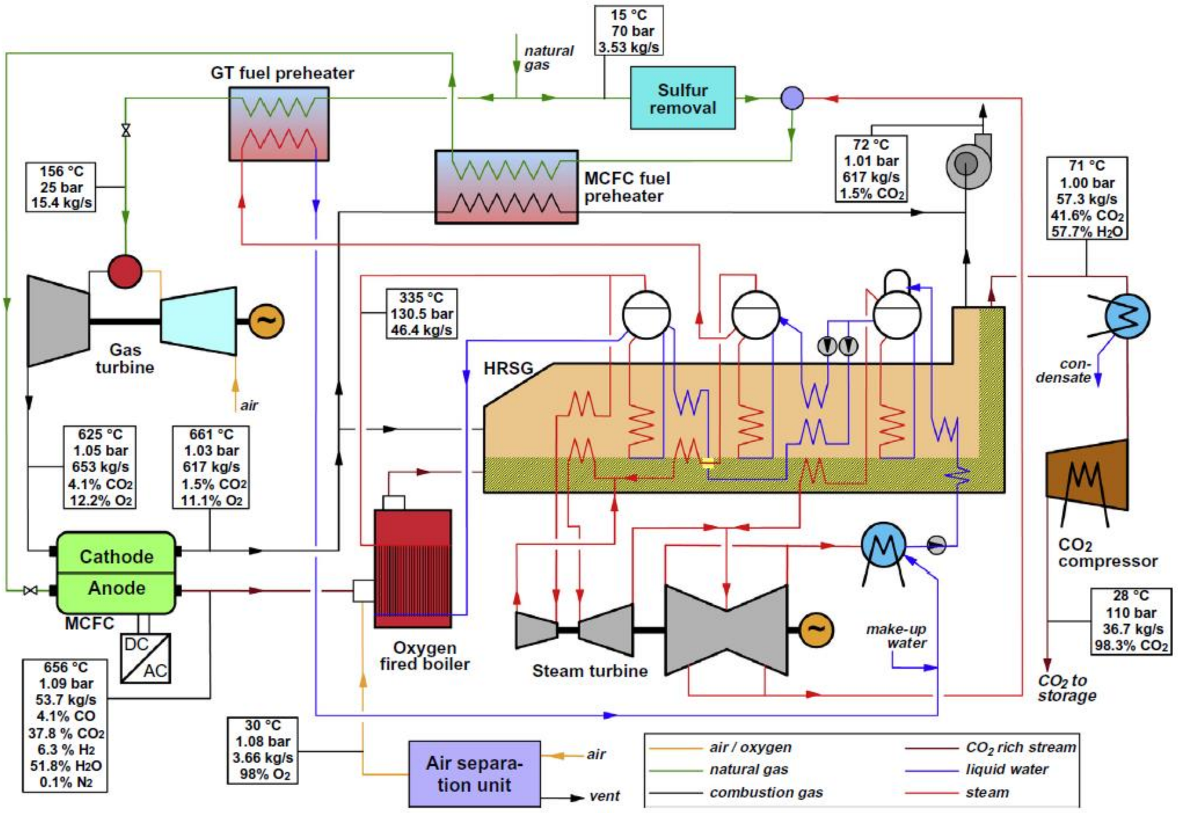


Figure 25: NGCC + MCFC plant configuration with oxy-combustion CO_2 capture (Oxy) (Campanari, et al., 2014)

Duan et al. (Duan, et al., 2015) presented two slightly different configurations from the above studies; in the first case a MCFC hybrid system with CO_2 capture integrated with both cryogenic separation and after-burner was considered.

The flowchart of this MCFC hybrid system is shown in Figure 26. Different from the previous systems, the cathode exhaust gas is split into two parts (splitter 3), one part is circulated to preheat the air and control the system's CO_2 capture ratio (mixer 2); the other part is sent to the HRSG to recover the waste heat and then vented to the atmosphere. The high purity oxygen produced from the cryogenic air separation unit is recycled to the after-burner to promote oxy-fuel combustion. The after-burner exhaust gas is cooled in the heater and then split into two parts: one part is circulated to the mixer 2, the other part is sent to the HRSG. The HRSG exhaust gas is a mixture of CO_2 and H_2O and as a result the CO_2 gas can be separated easily in a condenser and liquefied for storage after compression.

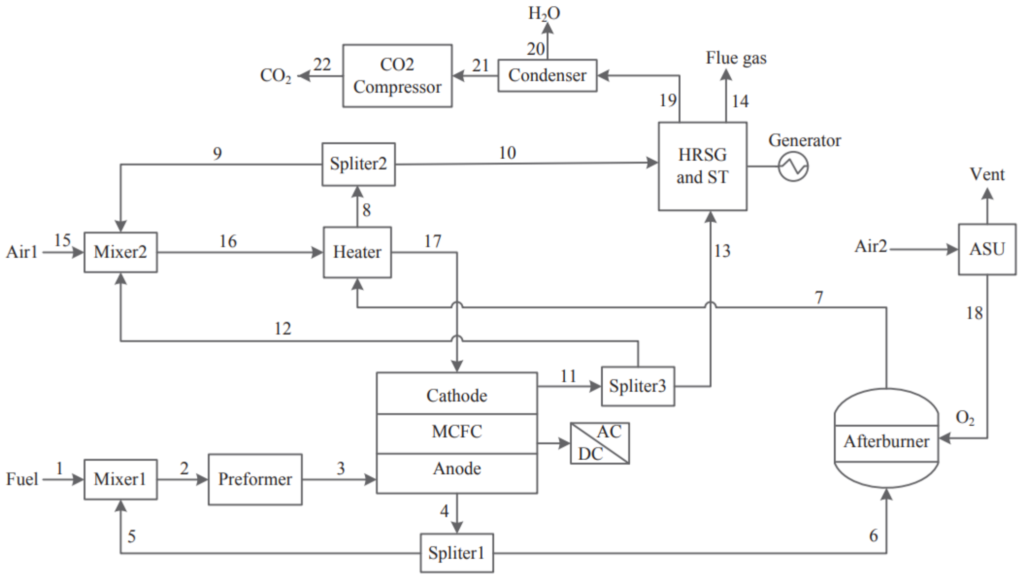


Figure 26: Plant flowchart of MCFC hybrid system with CO_2 capture integrated with the cryogenic air separation unit (Duan, et al., 2015)

In the second configuration, Duan et al. presented a MCFC hybrid system with CO_2 capture by integrating with the oxygen ion transfer membrane (ITM), as shown in Figure 27.

Similar to their previous case, the cathode exhaust gas is split into two parts. However, in this case the oxygen needed for the afterburner combustion is provided by an ITM unit. Air is compressed in the air compressor (Air 2) and heated to the required temperature for the ITM oxygen separation (Heater 1). The membrane separates the air into two parts: one part is the pure oxygen, the other part is the oxygen-depleted air at high pressure and temperature, the latter expands in an air turbine (AT) to produce power, and the expanded gas is sent to the Heater 3 to heat Air 1 and then vented to the atmosphere. In order to make full use of the waste heat, the Heater 4 is added before the condenser which utilises the heat of HRSG exhaust gas to preheat the Air 1.

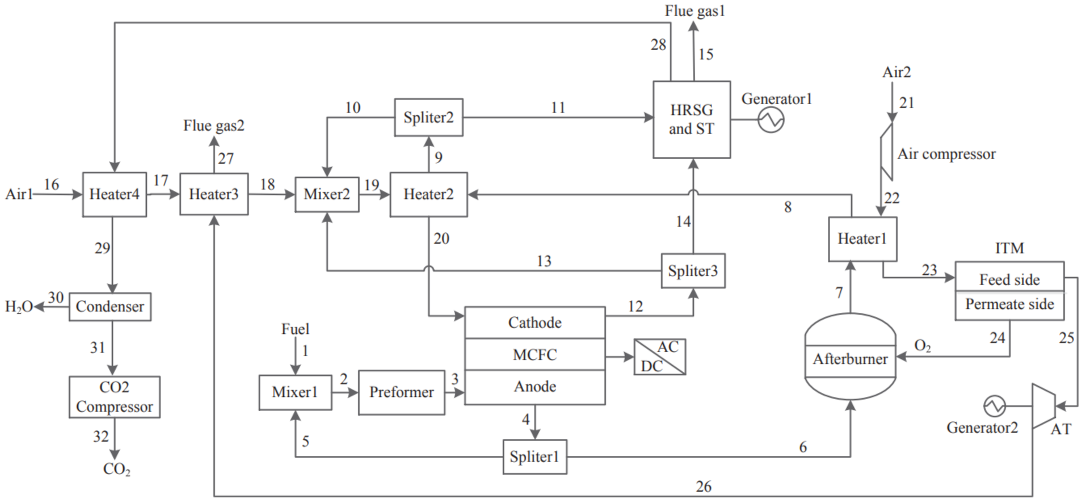


Figure 27: Plant flowchart of MCFC hybrid system with CO_2 capture integrated with ITM unit (Duan, et al., 2015)

The system net efficiency of the baseline system for which no CO_2 capture is applied is 63.36%, LHV. On the other hand, for the two systems with 85% CO_2 capture, the system's net efficiency is 60.94%, LHV and 62.68%, LHV for the cryogenic ASU and the ITM cases, respectively. The results of this study prove that the hybrid MCFC - ITM system is able to maintain high efficiency when capturing 85% of the CO_2 emissions, which is superior to the system integrated with cryogenic air separation unit.

Greppi et al. (Greppi, et al., 2013) studied an innovative configuration which included a NGCC – MCFC integrated with Gas Separation Membranes (GSMs) for CO_2 capture, taking into consideration operational constraints of both FCs and membrane technology. Integration of the NGCC power plant with a CO_2 concentration/separation MCFC-GSM unit is achieved by placing the MCFC-GSM unit between the open cycle (OC) and the steam cycle (SC) to treat the high-temperature OC exhaust, as shown in Figure 28. The OC exhaust (stream S05) is directed to the MCFC-GSM system to reduce its CO_2 content before being expelled as flue gas (stream S13), while the concentrated pure CO_2 stream (stream S12) is sent for transport and storage. The actual feedstock for the anode of the MCFC is also natural gas, provided by a different stream (stream S09) that is fed into the MCFC-GSM unit where it is internally converted to H_2 and then utilised in the anode.

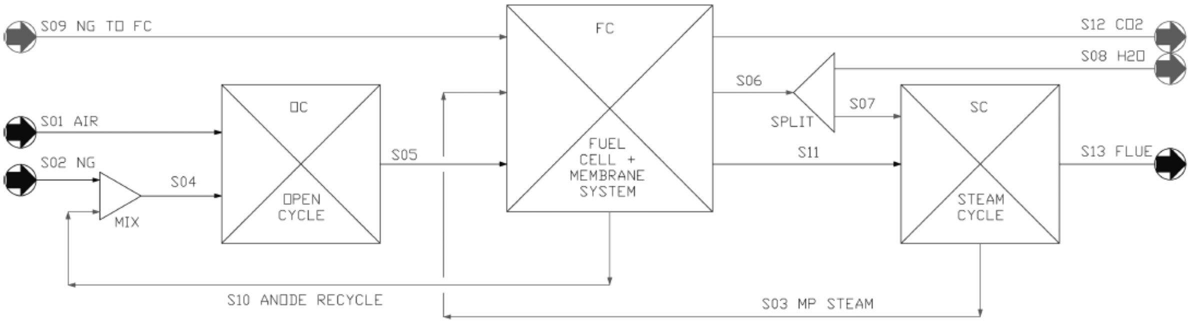


Figure 28: Process flow diagram for the NGCC-MCFC-GMS integration (Greppi, et al., 2013)

It should be noted that the natural gas fed to the MCFC anode needs to be desulphurised upstream, as seen in previous cases, in order to prevent contamination and/or carbon deposition within the MCFC unit. In this occasion the desulphurisation process also treats the natural gas fed to the OC (stream S02), as the same sulphur tolerance limitations exist for the OC exhaust stream (stream S05).

The medium pressure steam required for the fuel reforming process (stream S03) is obtained from the steam cycle, eliminating the requirement for a steam production subsystem. The heat requirements for the endothermic reforming reactions are covered by high temperature heat recovery from the pressurised burner in the Brayton cycle. The reasoning is that the natural gas burner can be adapted to deliver heat at the high thermal level required.

The proposed NGCC-MCFC-GSM system presented a 27% higher net electric output compared to an NGCC integrated with absorption (MEA) CO_2 capture system, due to the reduced energy penalty of the CO_2 separation from a concentrated stream. Another significant advantage is the additional power output gained from the MCFCs reaching an energy conversion efficiency of 52.5%, HHV (compared to 48.4%, HHV). However, the efficiency of the GSM CO_2 capture system reaches only 58.1%, whereas a MEA CO_2 capture system can reach 90% CO_2 removal. The same study also suggested that the COE for this configuration is between 70.0 and 72.3 €/MWh, while an MEA system lies between 78.6 and 79.1 €/MWh, depending on the CO_2 emissions price of the EU Emissions Trading System (ETS).

3.2.2. IGCC integrated with MCFC

This chapter focuses on the integration of an MCFC unit in IGCC power plants. One of the most important studies in this area was published by (Samanta & Ghosh, 2016). In their study an old operating 250 MW_e Indian coal-fired thermal power plant was considered.

The main modification for repowering the above plant was that a MCFC unit was placed downstream of the existing boiler, as seen in the schematic diagram of Figure 29. The MCFC's anode side is fed with a hydrogen-rich syngas generated from natural gas in an external reformer, while the cathode side is fed with the existing boiler exhaust of the IGCC plant after being desulphurised, diluted with fresh air and preheated. The reason given for diluting the boiler exhaust with fresh air is to effectively eliminate the possibility of harmful effects of traces gases before reaching the cathode of the MCFC.

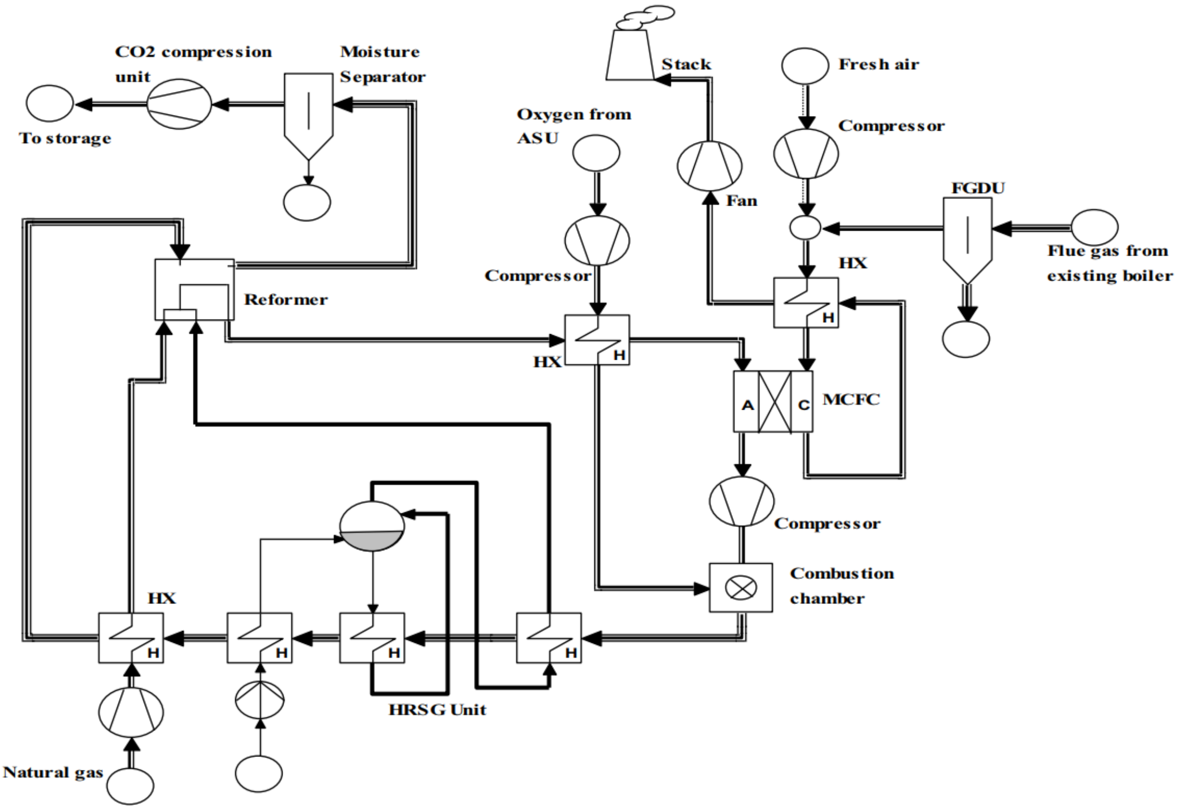


Figure 29: Schematic of the MCFC integration circuit in Cycle Tempo software interface (Samanta & Ghosh, 2016)

Most of the previous cases studied in this present report used internal reforming of the natural gas feed. However, this study considered an external reformer, utilising the heat of anode exhaust. The residual fuel from the anode exhaust is burnt with 98% pure oxygen supplied from an ASU followed by a HRSG unit, fuel preheating, heat supply to the external reformer, moisture separation and finally CO_2 compression and storage at 110 bar. The existing steam cycle configuration remains unchanged.

The proposed repowering configuration of the 250 MW_e IGCC power plant helps to increase net output and efficiency of the plant by about 27% and 1.1 percentage points (36.76% HHV, net efficiency), respectively, while capturing 67% of its CO_2 emissions. The COE of the repowered plant increases to 82 US$/MWh which is 46% higher compared to the COE of the existing plant and the cost of CO_2 avoided is 43.96 US$/tonne CO_2. Conventional CO_2 capture technologies (commercially available MEA) have reported lower capacities and efficiencies by 22% and 18%, respectively, when compared to the above proposed scheme, assuming the same amount of CO_2 captured. In addition, the COE of the IGCC plant with MEA CO_2 capture process increases to 90 US$/MWh, almost double than the COE reported for the proposed IGCC-MCFC process. These differences stem from the energy penalty incurred for the different CO_2 capture processes since the specific primary energy consumption for CO_2 avoided (SPECCA) of the proposed repowered plant is much lower (even negative) compared to the IGCC with MEA capture process. It is noted that an IGCC with MEA is not an obvious choice as a benchmark technology, however, this was the benchmark set in the 2016 study by Samanta & Ghosh. A separate IEAGHG study, “Further assessment of emerging CO_2 capture technologies”, is underway in which the new benchmark solutions will be presented.

Duan et al. (Duan, et al., 2016) scaled up Samanta's and Ghosh's design by modelling two scenarios; these incorporated a 251 MW_e and a 313 MW_e MCFC integrated in an IGCC cycle for 65% and 85% CO_2 utilisation (U_{CO2}) rates, respectively. This study reached a higher efficiency compared to Samanta's and Ghosh's study, reaching 47.31%, HHV, net, however, the COE was also higher, namely 103.85 US$/MWh. The innovative part of this study was the introduction of sensitivities with regard to U_{CO2} rate, U_f factor and current density (as previously discussed for the case of SOFC). The CO_2 capture rate is an important evaluation indicator of the new hybrid system, while the U_{CO2} rate is one of the important parameters that influence the above rates.

Figure 30 shows the net power output of a number of power plant sub-systems increases with increasing U_{CO2}. The fuel needed from the MCFC increases with the increase of U_{CO2}, hence, the exhaust gas mass flow of the anode side of the MCFC increases. The exhaust gas of the coal-fired power plant is constant, so the mass flow of CO_2 in the cathode is constant.

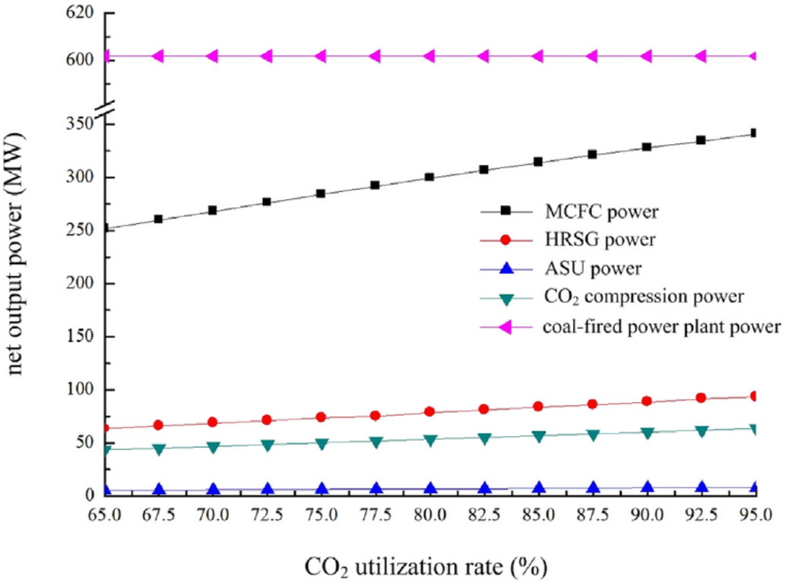


Figure 30: Effect of U_{CO2} rate on the power of various subsystems (Duan, et al., 2016)

The net power output increases because the ratio of the MCFC net output power to the net output power is higher and the rate of this ratio increases with an increase of U_{CO2}. However, the required input energy of the MCFC increases, i.e. the ratio of the fuel that the MCFC needs to the total fuel requirement increases, and the rate of change in fuel is greater than that of net total output power. So even if the system's total efficiency increases at first, it reaches a maximum around the U_{CO2} = 80% and then starts to decrease, as seen in Figure 31.

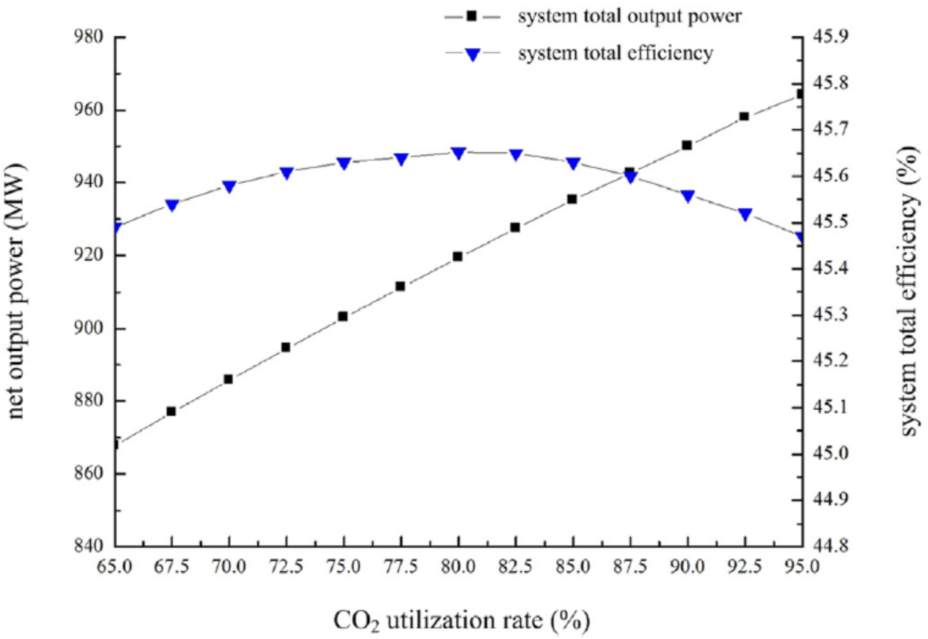


Figure 31: Effects of U_{CO2} rate on the system total output power and system total efficiency (Duan, et al., 2016)

Lastly, the CO_2 capture rate rises while the U_{CO2} increases, as seen in Figure 32. The CO_2 capture rate can rise to 96.21% when the U_{CO2} is 95%, however, it should be noted that these results represent simulations under ideal conditions. In reality the value of the U_{CO2} is difficult to reach 95%, a more representative value would be 80-85%, which corresponds to a CO_2 capture rate of 85-90%. In addition, a U_{CO2} of 95% would result in a significant efficiency reduction as seen previously in Figure 31. In order to reach a high total efficiency while maintaining a high CO_2 capture rate value and a high net output power, Duan et al. continued their sensitivity analysis keeping the U_{CO2} constant at 85%.

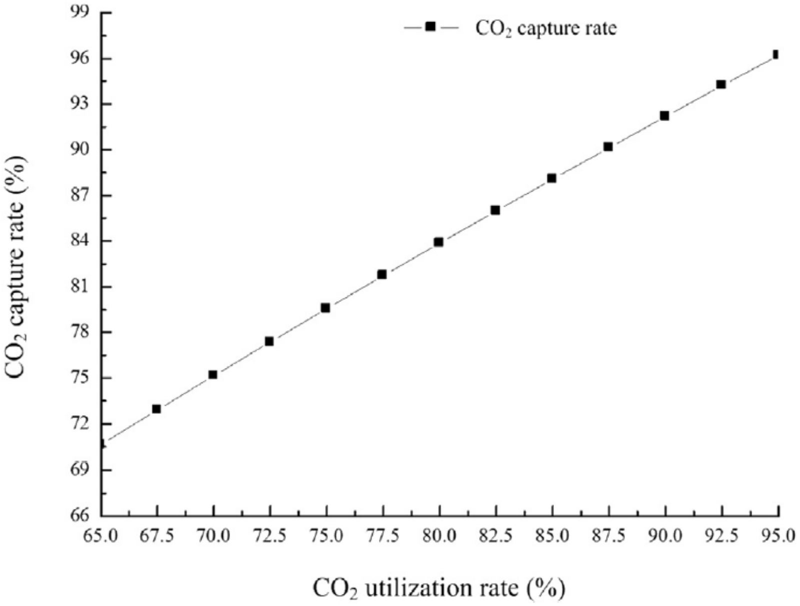


Figure 32: Effect of U_{CO2} rate on CO_2 capture rate (Duan, et al., 2016)

Since both the exhaust gas mass flow of the IGCC and the U_{CO2} are kept constant, the fuel that the MCFC anode consumes reduces with the increase of the U_f. As seen in Figure 33, both MCFC's current density and polarization losses increase which causes the decrease of the cell voltage. In order to obtain a high cell voltage, the U_f should be as low as possible, however, a low U_f means that the anode fuel is not consumed and in addition both the chemical energy loss of the fuel and the internal exergy loss of the cell increase. As a result it is safe to conclude that a U_f of 85% is a reasonable approach, being in accordance with most of the publications studied in this report as presented in Tables A and B in Appendix 1.

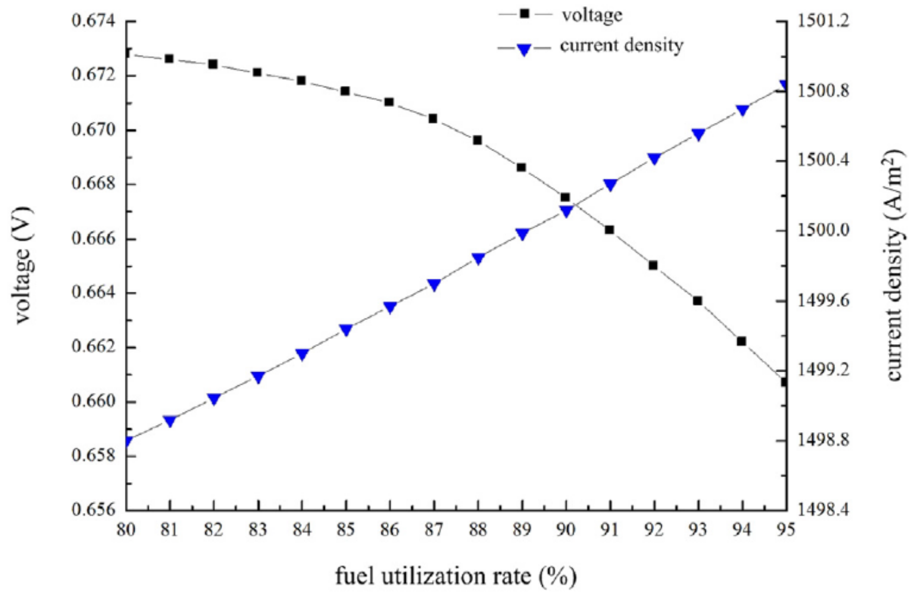


Figure 33: Effects of U_f rate on the MCFC voltage and current density (Duan, et al., 2016)

Figure 34 shows that for constant CO_2 mass flow and U_{CO2} in the cathode, the fuel that the MCFC anode consumes reduces with the increase of the U_f, so the H_2 provided by the fuel decreases which results in a decrease of the net output power of the MCFC. The effect of U_f on the mass flow of fuel is greater than that on the net output power of MCFC, so the system total efficiency rises firstly and then decreases. Since the H_2 provided by the fuel decreases, the CO_2 reacted in the anode decreases which causes the decrease of the CO_2 capture rate.

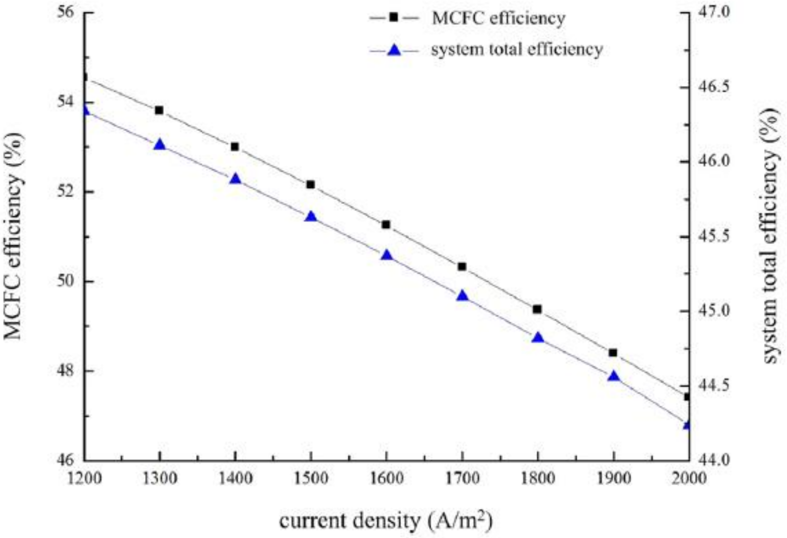


Figure 34: Effects of current density on the MCFC efficiency and system total efficiency (Duan, et al., 2016)

Regarding current density it was shown that when the U_{CO2}, U_f factor and the mass flow of exhaust gas of the IGCC are kept constant, the cell voltage decreases with the increase of current density, since the cell internal resistance increases and the reaction activity reduces.

As a result, the MCFC's net output power and efficiency as well as the system's total efficiency decrease, as seen in Figure 35.

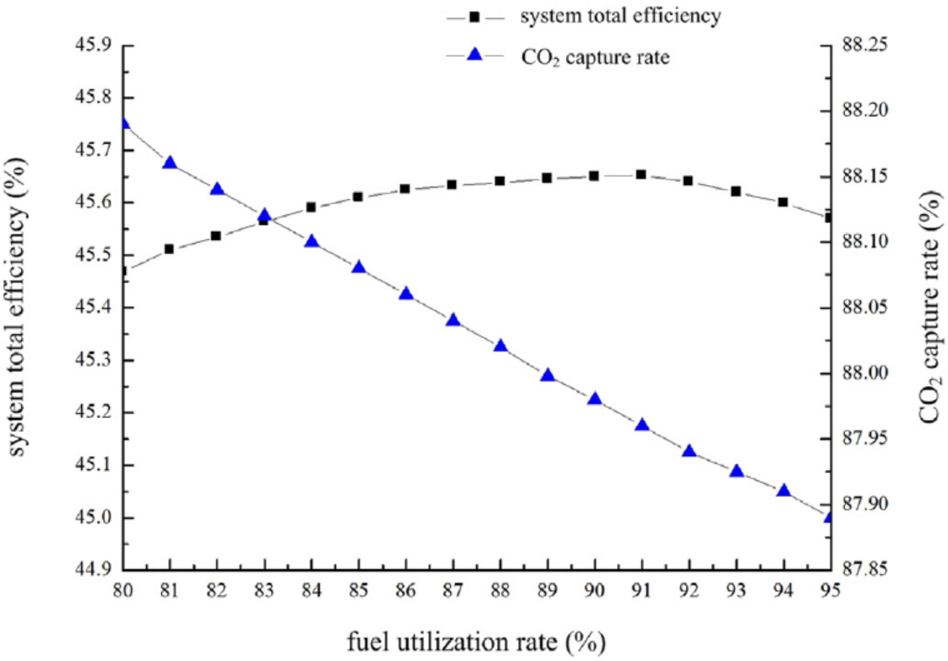


Figure 35: Effects of U_f rate on the system total efficiency and CO_2 capture rate (Duan, et al., 2016)

4. OPERATIONAL SOFC AND MCFC SYSTEMS

Currently power production from FCs installations are still scarce around the globe, however, there are signs of rapid development in the near future.

In the USA, California created a strategy to install large stationary FC plants which will provide a total output more than 100 MW_e, in order to meet their target of reducing Greenhouse Gases (GHGs). As a result, a number of large SOFC plants are in service such as Bloom Energy's 1 MW_e SOFC installation at Yahoo's Sunnyvale Headquarters in California. A FuelCell Energy 14.9 MW_e MCFC plant is in operation in Bridgeport, Connecticut.

In Korea, MCFCs of about 140 MW capacity have been installed by FuelCell Energy in collaboration with POSCO Energy, including the world's largest plant of 58.8 MW_e, installed in Hwasung City. In addition, Doosan has installed 35 Phosphoric Acid FCs (PAFC) in 2015 which have a total output of 15 MW_e. South Korea has also selected an 8.35 MW_e Bloom Energy SOFC plant for the Bundang Power Plant.

At the moment, multi-MW level SOFC and MCFC modules are manufactured by various companies and are being brought into the market. Operational experience data for a few selected case studies are now discussed.

4.1. The Siemens-Westinghouse system

The first successful demonstration of a SOFC-GT system was presented at the University of California in the U.S., using a Siemens/Westinghouse 220 kW_e tubular SOFC/GT hybrid with exhaust gas recuperation (Maurstad, et al., 2005). This system gave birth to the tubular SOFC designs that has been used quite frequently ever since.

Siemens – Westinghouse is a pioneer in tubular SOFCs, initially presenting a 100 kW_e SOFC-CHP system operating at atmospheric pressure in 1997. The system was able to produce 105 – 110 kW_e net AC power with an electrical efficiency of 46%, HHV. It was the first demonstration that utilised the commercial prototype cathode supported cells in stack reformers. The system operated for ~16,600 testing hours, becoming the longest-running FC in the world, at that time. The system consisted of 1,152 FCs with each cell having an 834 cm^2 effective area.

Siemens-Westinghouse proceeded to integrate a SOFC with a microturbine, creating a hybrid system, presented in Figure 36. The system operated at elevated pressure (3 atm) and it was able to produce 220 kW_e net AC power, 176 kW_e from the SOFC generator and 47 kW_e from the gas turbine (with 3 kW_e being lost for auxiliary loads). The system was fuelled with natural gas and achieved a net electrical efficiency of 57%. It operated for about 3400 hours from June 2001 to April 2002. This was the world's first SOFC-GT hybrid system.

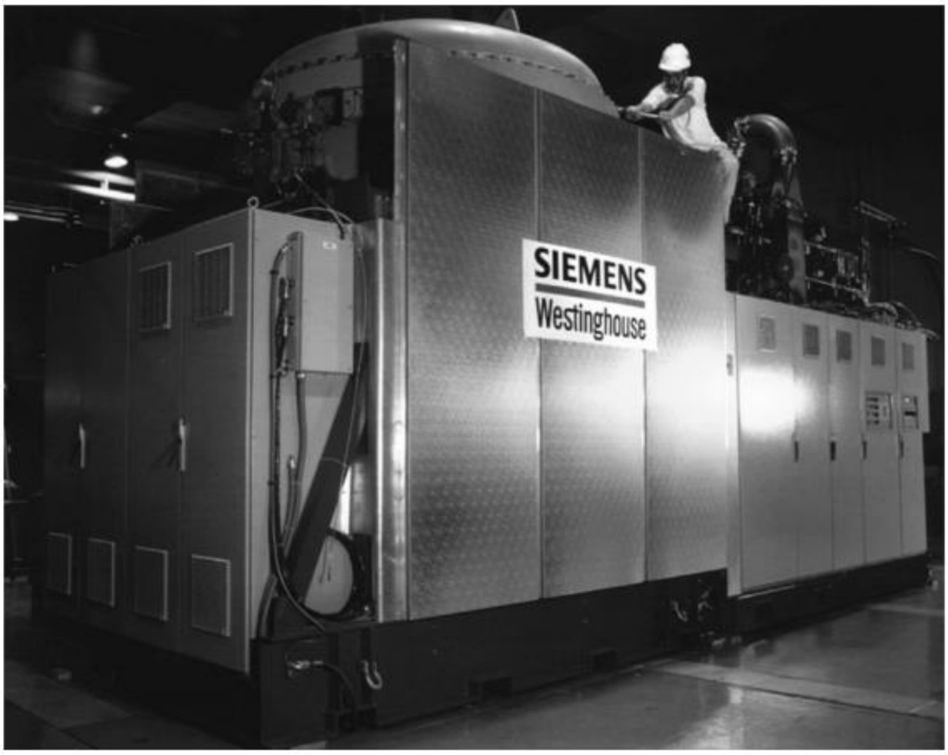


Figure 36: The hybrid SOFC – GT system developed by Siemens – Westinghouse (Maurstad, et al., 2005)

The operation of this hybrid system showed that coupling the SOFC with a microturbine is technically feasible but it requires further investigation regarding the microturbine design and integration. Specifically, the system required isolation valves in the main process loop to protect the FC system and helped to identify some of the challenges associated with effective control over transient conditions. Because SOFC technology is too expensive to risk using in the direct experimental study of system dynamics, SOFC / GT hybrid research has focused primarily on simulations for characterising transients and evaluating control strategies. To maximise both flexibility and fidelity of simulation results, a mixture of numerical models and small scale hardware designs has been used to explore the dynamics of SOFC / GT hybrids to varying degrees in at least three publicly available research facilities.

At the German Aerospace Centre (Deutsches Zentrum für Luft- and Raumfahrt e.V. - DLR) Institute of Combustion Technology, a 100 kW_e Turbec T100PH recuperated micro gas turbine was modified to include the piping required with the cathode loop of a SOFC and a valve to simulate the pressure losses associated with the representative FC and manifold system (Adams, et al., 2013). This facility can investigate compressor dynamics through thermal transients with parametric variation in FC pressure drop, arguably one of the most important parameters in component integration with turbomachinery.

Another hardware-based test facility was built by the Thermochemical Power Group at the University of Genoa. This also used a Turbec T100PH, but with the compressed airflow routed through a series of volumes, representative of the actual cathode and air manifold volume, connected by piping and isolated with valves (Pascenti, et al., 2007). The system can be used to parametrically vary the FC volume, facilitating the exploration of a number of different geometries. The University of Genoa experimental test rig has been used to validate a number of dynamic system models as well, based primarily on the Rolls-Royce Systems SOFC systems stack.

The U.S. Department of Energy designed and built a hardware-based simulation of a FC turbine hybrid power system at the National Energy Technology Laboratory (NETL), Morgantown, West Virginia Campus (Adams, et al., 2013). The facility utilises pressure vessels, piping and a natural gas burner controlled by a real-time FC model to simulate a high temperature FC. The hardware simulating the FC has been connected to a 120 kW_e Garrett Series 85 turbine and two primary surface exhaust gas recuperators for cathode preheat. Since the original combustion system of the Garrett turbine was removed, the FC simulator hardware in the NETL facility represents the only energy source to the system, and real-time (5 ms) coupling effects between the FC and turbine can be observed during simulations. Recently, a collaboration between NETL and the University of Genoa, a real-time gasifier model was added to expand capability of the facility to IGFC power systems (Traverso, et al., 2011).

As anticipated, hardware simulations have identified significant challenges in the transient operation of SOFC / GT hybrids. However, these have likewise identified various advantages apart from improved efficiency and provided solutions to some of the most critical operability issues, such as mitigation of compressor stall and surge during start-up (Adams, et al., 2013). In addition, the potential ability to turn down the total system power by up to 85% has been demonstrated for at least one system configuration at the NETL facility (Restrepo, et al., 2011). Numerical simulations of a pressurised, tubular SOFC / GT hybrid fuelled by natural gas have showed a negligible reduction in total system efficiency as FC U_f factor was decreased from 90% to 50%; as the U_f drops and the SOFC produces less power, the gas turbine produces more power correspondingly, resulting only a small change in the total power output. These results suggest that extensive FC degradation could be tolerated in a hybrid system while maintaining a constant power output without a substantial efficiency penalty. If higher FC degradation rates can be tolerated in a hybrid, the effective life of the FC would be extended, significantly improving the return on investment of any FC based power generation project.

4.2. The CEPACS System

FC Energy (FCE) has developed the Combined Electric Power and Carbon-dioxide Separation (CEPACS) system concept as a novel solution for greenhouse gas emission reduction (Ghezel-Ayagh, et al., 2017). The CEPACS system utilises electrochemical membrane (ECM) technology derived from the Company's well established Direct FuelCell® products (MCFC). A block flow diagram of the CEPACS system concept is shown in Figure 37 below.

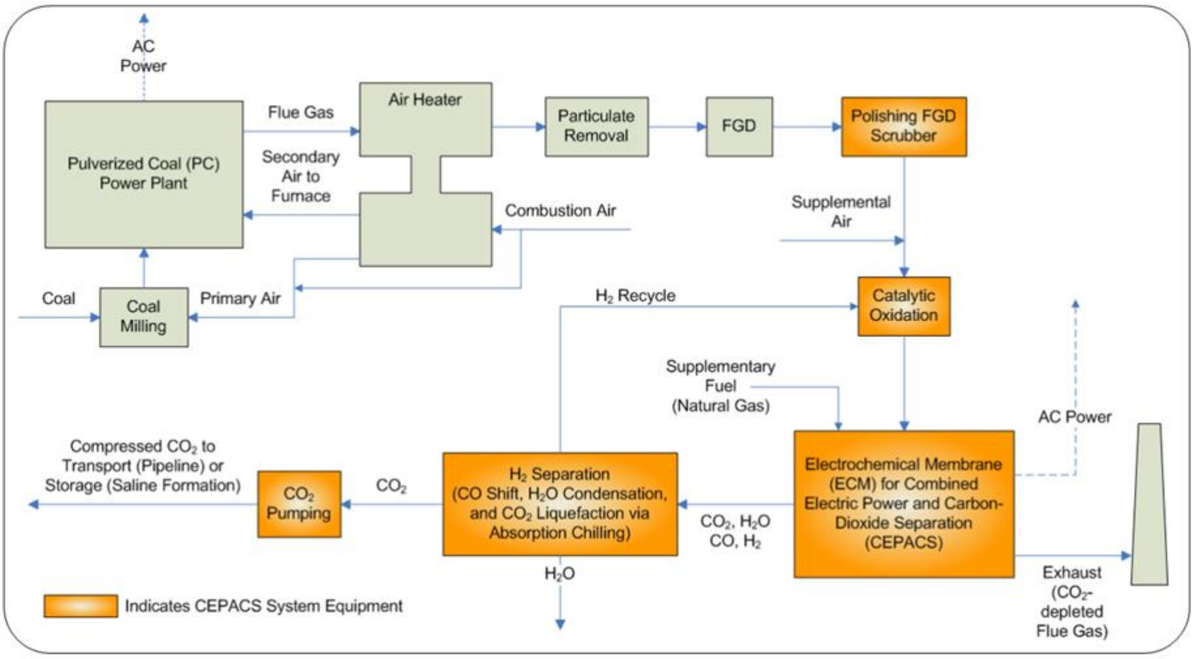


Figure 37: Block Flow Diagram of CEPACS System to Process Flue Gas from a 550 MW Pulverized Coal Plant (Ghezel-Ayagh, et al., 2017)

A. Process Concept

The flue gas from a pulverised coal (PC) power plant is sent to the CEPACS system after going through an existing flue gas desulphurisation (FGD) step. The flue gas is processed in a (secondary) polishing FGD scrubber to remove SO_2 to levels less than 1 ppm_v. The 1 ppm_v SO_2 limit was chosen based on a contaminant effects evaluation study for FCE's MCFC unit. Another study conducted by Chugoku Electric Power Company in Japan also targeted < 1ppm SO_2 in the feed to MCFC cathode, verifying FCE's results. After the polishing FGD scrubber the flue gas is pre-heated using FC waste heat from the cathode exhaust stream of the ECM. Additional air is introduced to the PC power plant flue gas (lean in O_2) to increase the O_2 concentration in the stream to a level suitable for ECM operation. The supplemental air is also preheated using the heat from the anode exhaust of the ECM. The mixture of PC flue gas/air stream is directed to a catalytic oxidiser. The fuel for the oxidiser is provided by a hydrogen-rich recycle stream. The H_2 is combusted in the catalytic oxidiser increasing the flue gas stream temperature to that required for the ECM operation. The hot stream exiting the oxidiser is then fed to the cathode-side of the ECM. Within the electrochemical membrane modules, similarly to the MCFC operational principles, CO_2 from the flue gas is transferred from the cathode side to the anode side. For each mole of CO_2 transferred, ½ mole of O_2 is also consumed in the cathode reaction.

The CO_2-depleted stream exiting the ECM cathode provides heat to preheat the natural gas fuel that is fed to the FC anodes and provides heat to generate steam (required for reforming of the natural gas fuel). The cathode exhaust stream also provides heat to the incoming flue gas stream, as discussed above. The CO_2-depleted cathode exhaust stream is then used to provide heat for a downstream multi-stage absorption chilling system before it is vented to the environment.

Regarding the anode side, natural gas is supplied as a supplementary fuel to the ECM anode resulting in simultaneous production of electric power and CO_2 separation. The natural gas is first desulphurised passing through an activated carbon fixed bed at ambient temperature. The activated carbon adsorption process effectively removes all organic (e.g. mercaptans) and inorganic (e.g. H_2S) sulphur compounds which could potentially damage the ECM anode electrode if untreated. Next, the desulphurised natural gas is mixed with steam and is then preheated as mentioned above. The fuel gas enters the ECM module where it is heated to 593°C within the fuel superheater (heat exchanger) that is located within the ECM module and is an integral component of the ECM stack. Methane is internally reformed in the ECM stack, producing H_2, which is the primary fuel for the anode. The carbonate ions formed in the cathode reaction migrate to the anode via an electrolyte in ECM cell and react with H_2 to form water vapour. The electrons released in the anode reaction flow to the cathode via an external electrical circuit which completes the power generation cycle. The DC electricity generated by ECM modules is converted to AC power using inverters.

The stream leaving the ECM anode contains unutilised fuel (mainly H_2 and CO) depending on the U_f, CO_2 transferred from the cathode and water produced during the FC electrochemical process. This stream is cooled down and directed to a shift converter where the CO is converted to additional CO_2 for capture. The stream leaving the shift converter flows to a condenser where water is recovered and is used in the process (includes water feed for steam required - NG reforming mentioned above). The dried stream then enters the first-stage absorption chiller, which utilises waste heat from the cathode exhaust stream, and is cooled to 2°C. After additional water separation, the cooled anode exhaust stream is compressed to 250 psia (~1.7 MPa) in multiple stages. In between the compression stages, the temperature is again lowered to 2°C using a combination of cooling water and subsequent stages of absorption chilling. After the final compression stage, a combination of the hydrogen-rich preheated recycled stream and the final stage chiller lowers the stream temperature to -44°C, at which point the CO_2 becomes a liquid as a supercritical fluid. At these conditions, however, H_2 remains in the gaseous phase and can be easily separated from the liquid CO_2 and this separation takes place in a flash drum. The H_2-rich stream is recycled back to the process to the catalytic oxidiser as well as to the anode fuel feed, as mentioned previously. The recycle of some H_2 to the ECM anode enables operation of the CEPACS system at high overall U_f reducing the natural gas fuel use and subsequently enhancing system's efficiency. The liquid CO_2 stream is then pumped to 2,215 psia (15,062 kPa) and heated using heat from the anode exhaust stream to 51°C for pipeline transport or sequestration.

The above CEPACS system employs 500 ECM modules (Ghezel-Ayagh, 2017). A standard module manufactured and used by FCE is shown in Figure 38. This module, called DFC1500, is capable of producing 1.2 MW_e of electricity. Each module consists of 4 stacks of carbonate FCs (ECM). Each stack has 401 cells.

Figure 38: Typical MW-scale FC Module by FCE (Ghezel-Ayagh, 2017)

In conventional power generation applications, MCFCs are operated at high power level and ~70% of the CO_2 in the cathode stream is transferred to the anode side.

As discussed previously, MCFCs performance is influenced by the partial pressures of the CO_2 and O_2 (reactants) in the feed stream and utilisation levels of the reactants in the anode and cathode. For CO_2 separation applications, the reactant conditions in the cell result in slightly lower electrochemical performance because of the higher than normal utilisation required to achieve greater than 90% CO_2 capture.

The cell performance during operation in a CO_2 transfer mode was determined in laboratory scale testing. Cell voltage was measured as a function of the CO_2 transfer rate or flux defined as cc/s/cm^2. Figure 39 shows the resulting performance for three levels of U_{CO2}, (namely 70%, 80% and 90%). The CEPACS design with 500 ECM modules removing 90% of the CO_2 from the PC plant flue gas has a CO_2 flux rate of about 0.0135 cc /s/cm^2 which suggests a cell performance level of about 750 mV based on the lab scale testing presented. Design of the CEPACS system is based on a performance of 790 mV, based on recent improvements in cell performance and operation at lower than normal current density (power) level of 110mA/cm^2.

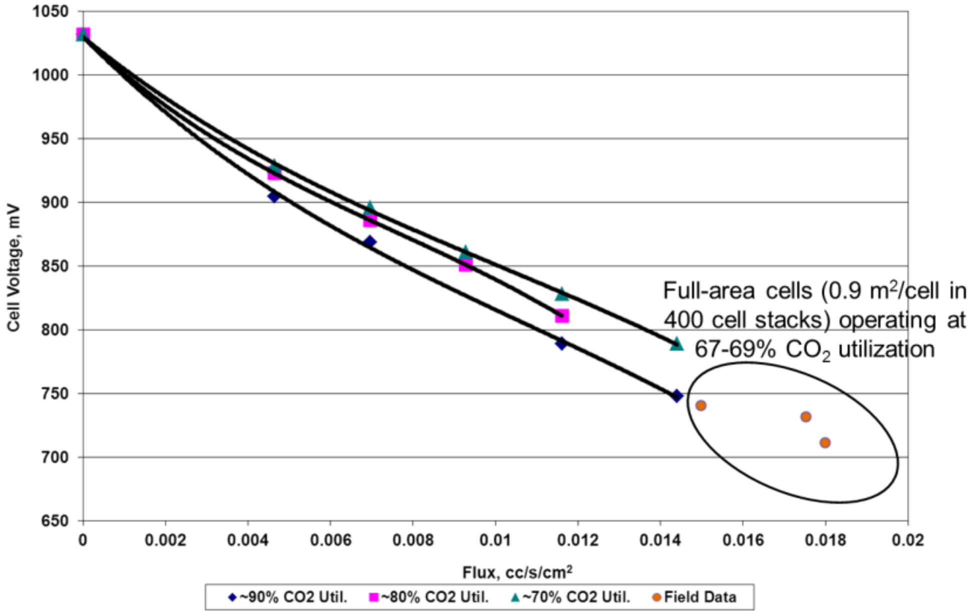


Figure 39: Preliminary Cell Performance Based on Subscale Cell Tests (Ghezel-Ayagh, 2017)

B. Effect of Contaminants

Next, contaminant tolerance levels for the ECM were identified, based on experimental results. The contaminant levels expected after the flue gas clean-up sub-system were estimated and compared with the ECM tolerance levels, seen in Table 4. It is clear that the ECM tolerance levels are well above the contaminant levels expected in the ECM cathode feed gas (treated flue gas).

Table 4: Contaminants levels and their effects (Ghezel-Ayagh, 2017)

Flue Gas Contaminant	Highest Concentration Tested, with Negligible Power Degradation		Estimated Concentration in Cathode Inlet Gas after Polishing FGD		Notes
SO_2	1	ppmv	0.18	ppmv	Performance losses due to short term SO_2 exposure up to 40ppm were fully reversible
Se	10	ppbv	0.3	ppbv	No apparent degradation over 860 hours.
Hg	250	ppbv	0.08	ppbv	Expected form is predominantly elemental Hg. No apparent degradation over 1100 hours.
HCl	200	ppbv	12.7	ppbv	No apparent degradation over 900 hours.

C. CEPACS Test Results

The U_f and U_{CO2} as well as the current densities used as the operating points for the testing period are listed in Table 5. Five utilisation conditions in the first column along with the current densities at each flow condition constituted the test matrix. The 100% flow condition corresponds to the oxidant and fuel flow rates required to achieve 92% U_{CO2} and 68% U_f, respectively at 110 mA/cm^2.

Table 5: Experimental testing using various U_f and U_{CO2} (Ghezel-Ayagh, 2017)

	Current Density			
	67% Flow	**90% Flow**	**112% Flow**	**135% Flow**
	mA/cm^2	**mA/cm^2**	**mA/cm^2**	**mA/cm^2**
30% U_f & 40% U_{CO2}	29.6	40.4	51.6	62.4
43% U_f & 59% U_{CO2}	44.8	60.8	77.2	93.2
52% U_f & 71% U_{CO2}	54.8	74.4	94.0	113.6
59% U_f & 80% U_{CO2}	62.4	84.4	106.8	128.8
68% U_f & 92% U_{CO2}	72.8	98.4	123.6	149.2

The ECM FC test results associated with variations to operating temperature and cathode inlet oxygen concentration conditions are shown in Figure 40 and Figure 41, respectively.

The bar chart in Figure 40 shows the effect of cell operating temperature on cell performance at cathode inlet oxygen concentration of 8.3% v/v, wet for various reactant utilisations, for the highest flow condition of 135% flow. Based on Table 5, it can be seen that the comparison at 68% U_f and 92% U_{CO2} condition (right side of the bar chart) corresponds to the cell performance at 149.2 mA/cm^2. Whereas the comparison at 52% Uf and 71% U_{CO2} condition (middle of the bar chart) corresponds to the cell performance at 113.6 mA/cm^2. Increase in cathode inlet O_2 concentration represents the extent of air supplementation of the flue gas. As the O_2 concentration is enhanced by the air addition, the CO_2 concentration decreases.

The bar chart in Figure 41 shows the effect of cathode inlet oxygen concentration on cell performance at 620°C operating temperature for various reactant utilizations, for 112% flow condition. The comparison at 68% U_f and 92% U_{CO2} corresponds to the cell performance at 123.6 mA/cm^2. Whereas the comparison at 59% U_f and 80% U_{CO2} condition corresponds to the cell performance at 106.8 mA/cm^2.

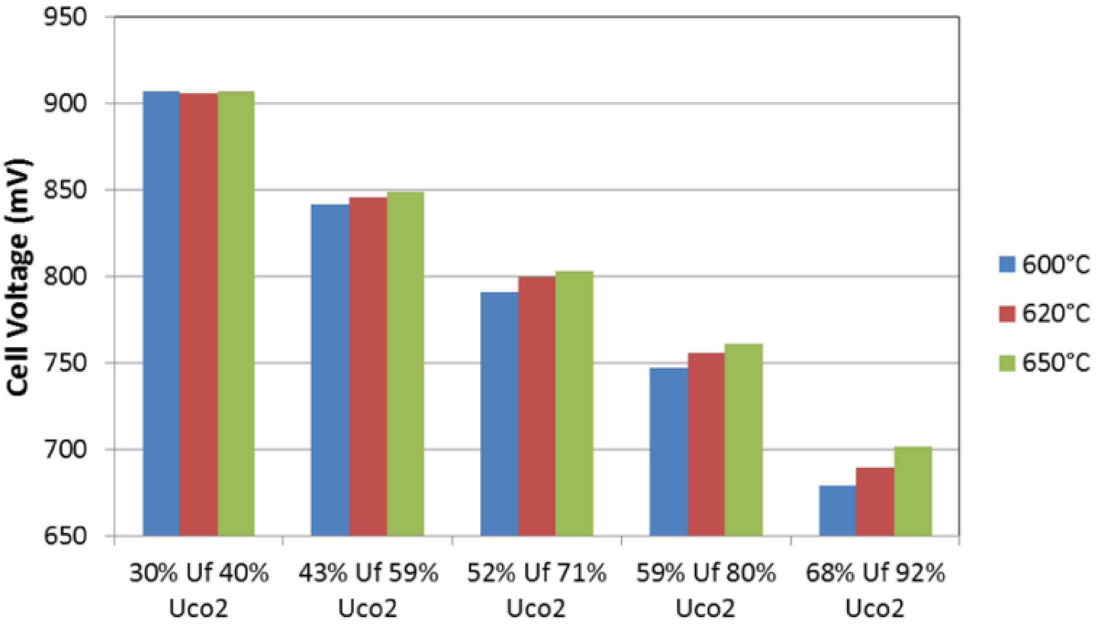


Figure 40: Effect of Operating Temperature on ECM Cell Performance with 8.3% O_2 (9.9% CO_2) Cathode Inlet Gas Concentration (135% fuel & oxidant flows) (Ghezel-Ayagh, 2017)

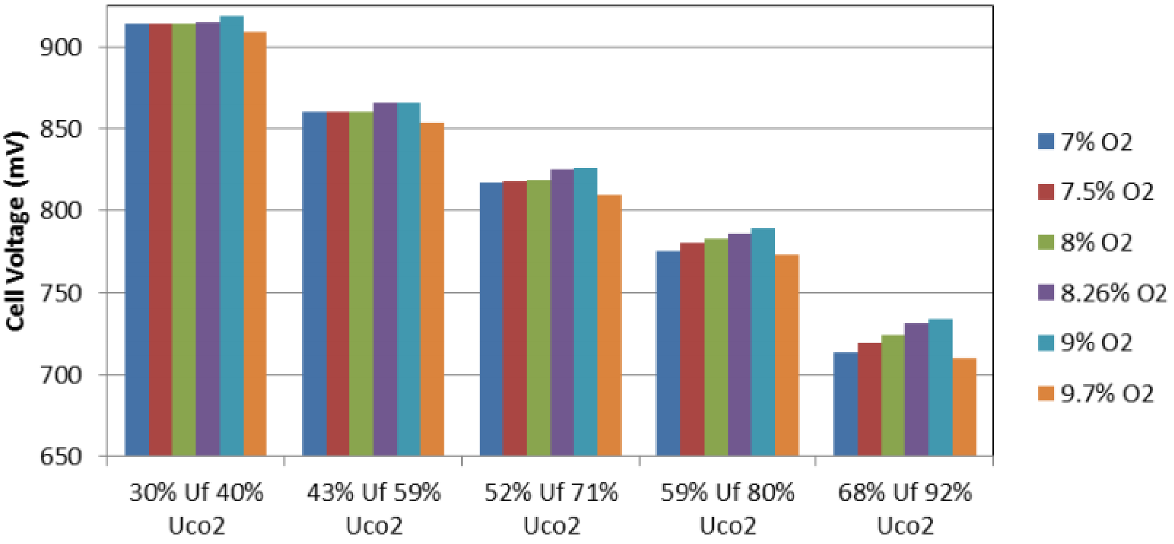


Figure 41: Effect of Cathode Inlet O_2 Concentration (air supplementation of flue gas) on ECM Cell Performance at 620°C Operating Temperature (112% fuel & oxidant flows) (Ghezel-Ayagh, 2017)

The above charts indicate that the cell performance increases with the cell operating temperature. It is also apparent that as the cathode inlet oxygen concentration increases up to 9.0% v/v, wet, the cell performance increases. However, the performance decreases once the oxygen percentage reaches 9.7% v/v, wet (performance loss due to corresponding lower cathode inlet CO_2 on centration may be dominating). The resultant cell performance was affected increasingly by the operating temperature, cathode inlet oxygen concentration, and flow variations as the U_f and U_{CO2} increased.

D. Cost of Electricity and CO_2 Captured

A unique feature of the CEPACS system is that in the process of capturing ≥90% of the CO_2 from the 550 MW_e (net AC) PC plant flue gas, it generates 351 MW_e additional (net AC) power at the same time after compensating for the auxiliary power requirements of CO_2 capture and compression.

The net electrical efficiency of the CEPACS-equipped PC plant including CO_2 capture was estimated to be 38.8% (based on HHV of coal and natural gas for the PC plant and CEPACS system, respectively). The Total Operational Cost (TOC) of the subcritical PC power plant with the CEPACS system was estimated to be 2,297 US$/kW and the estimated Cost of Electricity (COE) is 80.9 US$/kWh, 25% lower compared to NETL's case of CO_2 capture using amine solvents in a PC plant. The cost of CO_2 captured was calculated for those two CO_2 capture cases of (i) Case 2 PC with Amine CO_2 capture; and (ii) Case 3 PC + CEPACS. The costs are shown in year 2007 US dollars and per tonne of CO_2, for consistency with the Technical and Economic Feasibility Study (T&EFS) basis. Figure 42 shows the Cost of Electricity (COE) with component breakdown and Incremental COE (relative to base Case 1 PC without CO_2 capture) calculated for each case.

The following key conclusion can be drawn:-

- The CEPACS system results in a low COE, offering CO_2 capture at low additional cost. The incremental COE for the CEPACS is 36.2%, significantly lower compared to the amine CO_2 capture system that is 84.3%.

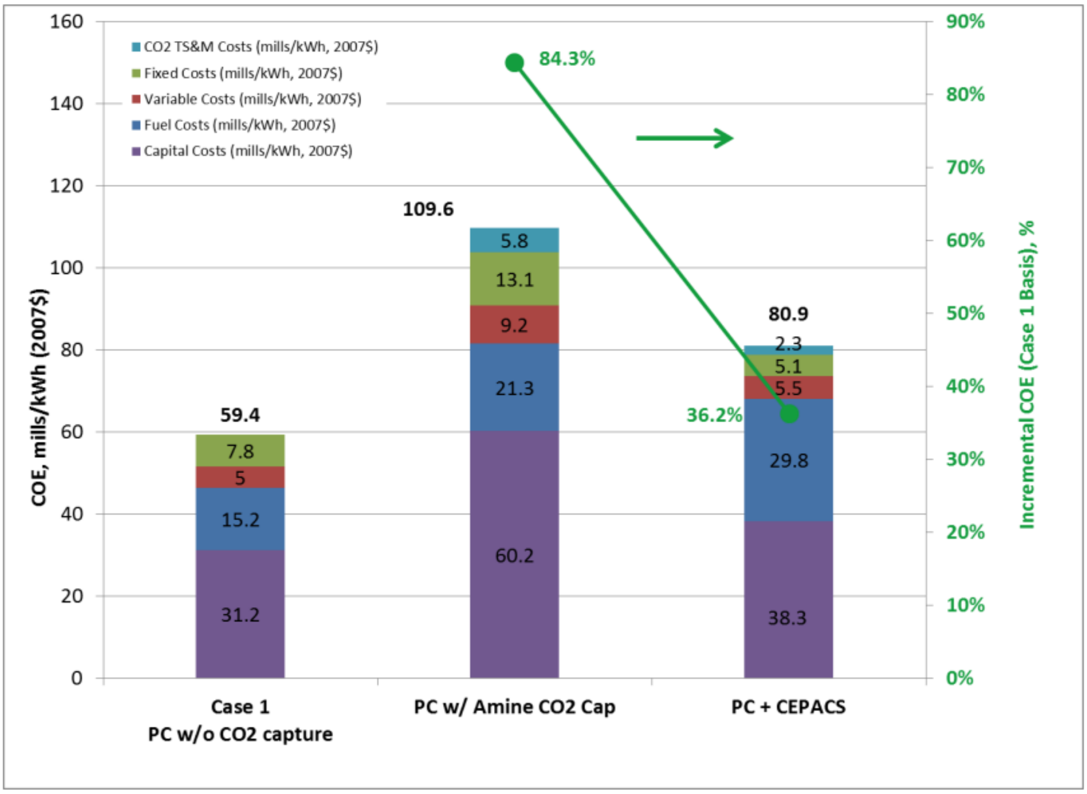


Figure 42: Breakdown of Cost of Electricity (COE) and Incremental COE (Ghezel-Ayagh, 2017)

Figure 43 shows the cost of CO_2 captured calculated for the two CO_2 capture cases of (i) Case 2 PC with Amine CO_2 capture; and (ii) Case 3 PC + CEPACS. The costs are shown in year 2007 US dollars and per tonne of CO_2, for consistency with the T&EFS estimation basis. The following key conclusion can be drawn:

- The CEPACS system cost of CO_2 captured is 33.63 US$/tonne (in 2007 USD), which compares favourably with the US DOE target of less than 40 US$/tonne (2011 USD).

- The cost of CO_2 captured for the CEPACS system is 27.2% lower than that for the amine CO_2 capture process (Case 2).
- The cost of CO_2 avoided for the CEPACS system is 60.3% lower than that of the amine system, due to the extra power generation of the CEPACS plant which results in significantly lower CO_2 emissions (lb/MWh), at a fixed 90% CO_2 capture rate.

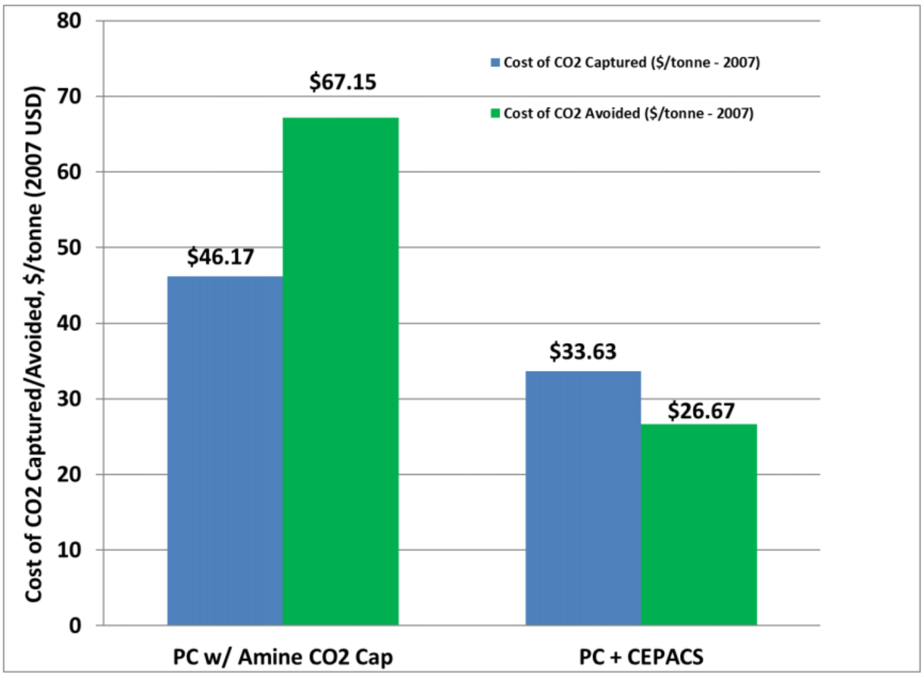


Figure 43: Cost of CO_2 Captured/Avoided (Ghezel-Ayagh, 2017)

4.3. The BioZEG System

BioZEG (Biomass Zero Emission Gas) power system is configured on a hybrid technology for co-production of hydrogen and electricity from hydrocarbon feedstock integrated with a CO_2 capture system (Andresen, 2015).

The developers of this technology claim that a plant efficiency of >80%, LHV can be reached, including almost 100% CO_2 capture and compression to 110 bar. The co-production of electricity and hydrogen is achieved by using a Sorption Enhanced Reforming process (SER or SE-SMR, sorption enhanced steam methane reforming) for hydrogen production and SOFC technology for electricity production. The heat produced in the SOFC is used to regenerate the CO_2 sorbent, eliminating the need for an after-burner. Excess hydrogen from this process is collected as a product. As a result, integration of CO_2 capture is achieved without an energy penalty, thus reducing the investment costs, fuel consumption, while eliminating NO_X emissions and increasing overall efficiency. The produced hydrogen is then fed to the FC stacks, increasing the power density in the stacks and reducing the investment cost of the FC system. In addition, the modular BioZEG system offers significant flexibility benefits with regards to applications and scale, from small scale distributed plants to industrial scale power plants. By using biomass as a feedstock the technology provides a carbon negative solution.

Sorption enhanced reforming (SER) technology for hydrogen production has been studied extensively and the concept of integrating a CO_2- capture is currently emerging. In this process both the reforming and CO_2 capture steps are integrated within two reactors, enabling hydrogen to be produced in one single step. The water gas shift (WGS) section is eliminated. A CO_2 sorbent, in this case CaO, is mixed with a reforming catalyst, resulting in the removal of the CO_2 in the synthesis gas mixture as it is formed. This causes the reforming and water gas shift reactions to proceed simultaneously beyond the conventional thermodynamic limits. In addition, capturing CO_2 ‘in situ’ enables the recovery of high purity CO_2 after sorbent regeneration, eliminating costly downstream separation steps.

The main advantage of the SER system compared to conventional methane reforming is the simplicity of the process. Reforming, water gas shift and CO_2 separation occur simultaneously in the same reactor, a higher hydrogen yield is obtained in the single step reaction (95-99 mole% dry basis), reducing the need for downstream H_2-purification while potentially increasing efficiency, energy savings and reducing production costs.

A. Process Description

The BioZEG-plant is a 50 kW_e prototype that was built and installed at Hynor Lillestrøm, a renewable hydrogen station and technology test centre in the city of Lillestrøm, Norway (Andresen, 2015). The main objective of the test centre is to design, build and operate a hydrogen station based on local renewable energy sources and to demonstrate new and innovative hydrogen technologies.

Hynor Lillestrøm is equipped with a pipeline from a municipal waste landfill site and an upgrading gas system providing a bio-methane output of 85-90% purity. The BioZEG-plant consists of a 30 kW_{H2} SER reactor system, a 20 kW_e SOFC module and a high temperature heat exchange section, for close thermal integration of the SER and SOFC systems. The plant is constructed in such a way that both SER and SOFC can be tested and optimised separately for stand-alone production of hydrogen and electrical power.

The system layout of the BioZEG plant can be seen in Figure 44.

BioZEG system is divided into four sub-systems: (i) a SE-SMR reactor system, (ii) a SOFC-module, (iii) thermal system integration of the two basic technologies and (iv) balance-of-plant (BOP).

The SER reactor system is a dual bubbling fluidised bed (DBFB) reactor system consisting of one reformer and one regenerator. The SER-reformer has a hydrogen production capacity of about 10 Nm^3/hour (1 kg/hour). The reformer is operated at around 600°C and bio-methane is reformed in the presence of steam and a Ni-based catalyst. CO_2 is simultaneously captured by a solid CaO-based sorbent, arctic dolomite. When the SER reactor system is operated as a part of the BioZEG-plant the temperature in the regenerator is increased to around 850°C by surplus heat from the SOFC-system causing the release of CO_2 into a separate gas stream. Thus, the SER reactor system produces close to pure hydrogen from bio-methane while separating the CO_2. For optimal performance of the ZEG system, high temperature SOFC (1000°C) with ceramic interconnects is preferred due to the high temperature needed for regeneration of the solid sorbent. However, in this plant a medium temperature SOFC (850°C) with metallic interconnects has been chosen in order to reduce overall project risks and complexity. Compared to an efficiency optimised configuration of a ZEG plant, three adjustments were made; the SOFC operating temperature is reduced, SOFC fuel recycling is excluded, and an after-burner is added increasing the temperature.

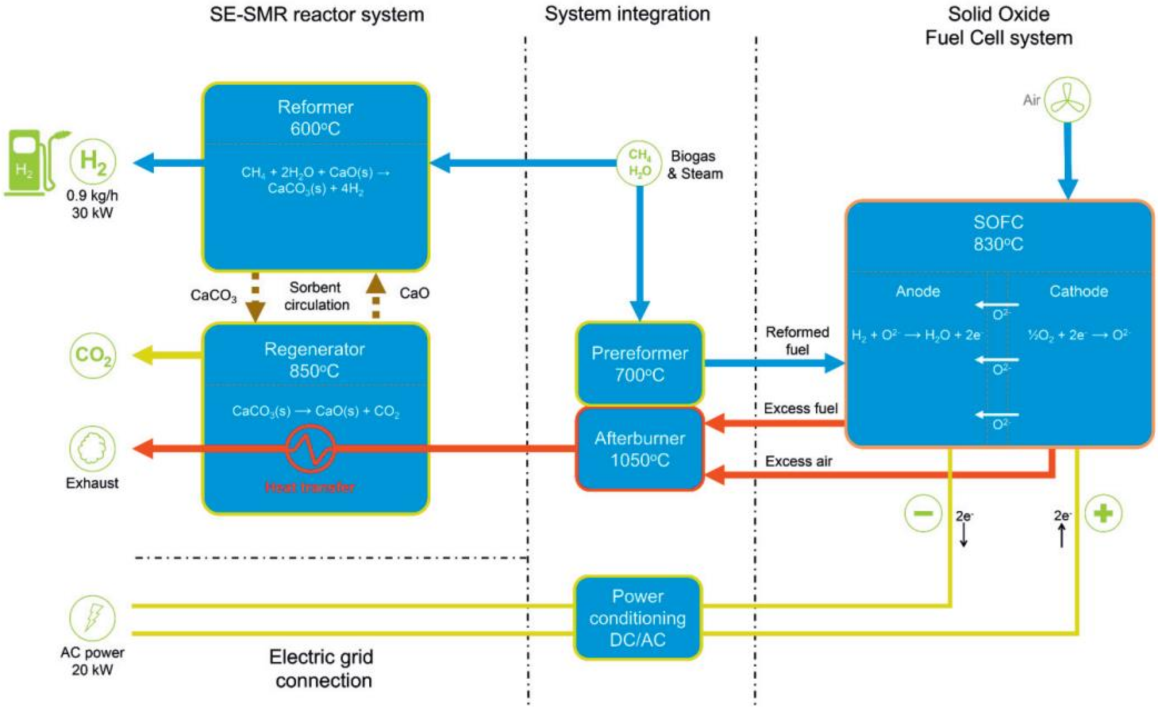


Figure 44: Schematic BioZEG-plant system layout (Andresen, 2015)

The SOFC module is custom made for the BioZEG-plant, engineered and assembled by CMR Prototech. The SOFC-module consists of 24 SOFC-stacks each made of 30 cell plates (130 x 150 mm) with metallic CFY (chromium-iron-yttrium) interconnects. The stacks were delivered by a European consortium led by Plansee and Fraunhofer IKTS.

Regarding the SOFC system, a dual stack-configuration is chosen and the module consists of 12 hot boxes with two stacks in each box. At a nominal operating temperature of 810-840°C, the SOFC module has a rated power capacity of 20 kW_e running on previously reformed bio-methane. The biogas that is fed to the anode of the SOFC is mixed with steam before entering a pre-reformer. This unit is integrated with an after-burner and the combined pre-reformer / after-burner unit produces the required fuel quality for the SOFC module and increases the temperature up to the requirement of the SER regenerator. High temperature gas-to-gas heat exchangers and other core components were also developed during the project.

B. Initial Test Results

Long term operational stability and low degradation rates of the SOFC module are important for industrial commercialisation of the BioZEG technology.

As a result, the most crucial parts of the design were thoroughly tested prior to the commissioning and operation of the pilot plant. The baseline performance of a single dual stack-box was tested and verified, operated both at fixed load conditions and at varying load levels for more than 1000 hours. The test period also included several thermal cycles. The dual stack performance during one thermal cycle is shown in Figure 45.

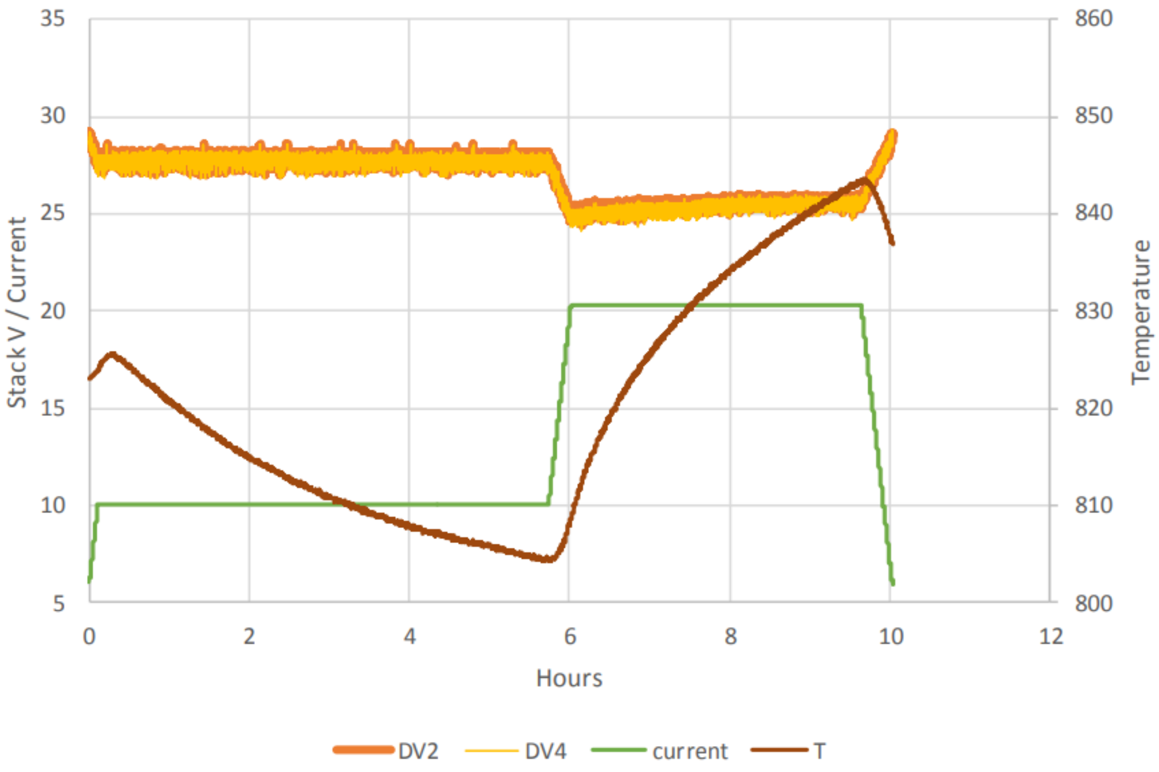


Figure 45: Dual stack-box performance during one thermal cycle. DV2: Voltage over stack no 1; DV4: Voltage over stack no 2; T: temperature (C) (Andresen, 2015).

The anode side fuel feed was a mixture of hydrogen (55%), nitrogen (32%) and steam (13%) at a rate of 23 Nl/min. The air flow rate to the cathode side was 200 Nl/min. At 20 A this corresponds to a U_f factor of 66% at which the power production was around 1 kW_e. Initial tests showed that the electrolyte supported cells can tolerate several redox cycles without significant degradation. The thermal expansion of the materials is well matched and the stack can be cooled and heated without observable loss in performance. High thermal mass means that significant heat is required for start-up, however, start-up can be still performed in less than 6 hours.

The main challenge of the integrated operation of the BioZEG concept is the integration of the heat transfer from the SOFC module to the SER regenerator. The thermal integration is crucial in order to achieve high system efficiencies. To the best of the authors' knowledge this challenge still remains and there have been no significant advancements in this research. Verification of the feasibility of the designed solution for thermal integration and optimisation of the operating parameters of the heat exchangers is still ongoing. It is anticipated that depending on the application of the BioZEG concept, the system will be configured for different operating conditions with regards to load variations, load following rates, idling and peak load demands in various configurations. Furthermore, another parameter under investigation would be the fuel composition, especially when using biomass based fuels such as biogas. Different steady-state modes are planned for testing in order to determine peak load limits and the transients and load programs will be based on system demands in different industrially integrated processes.

5. TECHNO-ECONOMIC ANALYSIS

This section of the report aims to present an economic feasibility assessment of a number of FC hybrid cycle arrangements. In each case the input data has been rationalised and homogenised to ensure that a fair comparison can be made. The selected FC hybrid cycle arrangements have also been compared against a number of reference cases for competing technologies.

5.1. Case Descriptions

The cases investigated are as follows:

Case	Description
Reference Case 1a	Supercritical Pulverised Coal (SC PC) boiler plant without CCS (NETL, 2015)
Reference Case 1b	Supercritical Pulverised Coal (SC PC) boiler plant with CCS (Cansolv CO_2 Capture Process) (NETL, 2015)
Reference Case 2a	Natural Gas Combined Cycle (NGCC) plant without CCS (NETL, 2015)
Reference Case 2b	Natural Gas Combined Cycle (NGCC) plant with CCS (Cansolv CO_2 Capture Process) (NETL, 2015)
Reference Case 3a	Integrated Gasification Combined Cycle (IGCC) plant (GEE Radiant Gasifier) without CCS (NETL, 2015)
Reference Case 3b	Integrated Gasification Combined Cycle (IGCC) plant (GEE Radiant Gasifier) with CCS (2-Stage Selexol CO_2 Capture Process) (NETL, 2015)
1	Atmospheric integrated NGFC (utilising MCFC) with CO_2 purification through cryogenic CO_2 separation. (Campanari, et al., 2014)
2	Atmospheric IGFC (utilising SOFC) with CO_2 capture through oxy-combustion followed by condensation. (NETL, 2009)
3	Pressurised IGFC (utilising SOFC) with CO_2 capture through oxidation followed by flash cascade. (Adams & Barton, 2010)
4	Pressurised integrated NGFC (utilising SOFC) with CO_2 purification through oxidation followed by flash cascade. (Adams & Barton, 2010)
5	Atmospheric MCFC with Natural Gas at the anode and exhaust from SCPC boiler at the cathode, with CO_2 capture through oxy-combustion followed by condensation. (Duan, et al., 2016)

A summary of the key parameters for all the cases considered in the evaluation are shown in Appendix 3 Table A3-1.

5.2. Evaluation Basis

For each case the ‘bottom-up’ capital cost assessments were taken from the available data obtained in the literature review. The breakdown of capital costs for key plant items was taken and collated along with any assumptions used in the studies, such as currency, plant location and cost year.

The capital costs used are on a “Total Plant Cost” (TPC), which comprises the Bare Erected Cost (BEC) plus the cost of services provided by the engineering, procurement and construction (EPC) contractor (such as detailed design, permitting and project management) and project and process contingencies.

The BEC comprises the cost of process equipment, on-site facilities and infrastructure that support the plant (e.g. shops, offices, labs, road), and the direct and indirect labour required for its construction and/or installation.

For the purposes of rationalisation, the breakdown of the TPC for each individual plant item was assumed to be as follows:

BEC

- 60% Equipment Cost
- 30% Labour Cost
- 10% Materials Cost

TPC

- Contingency 10% of BEC
- OH and Management 7.5% of BEC

Limitations in the techno-economic analysis are noted. None of the technologies considered here have been fully commercialised and the techno-economic data obtained from the literature review is generally based on theoretical performance predictions from modelling and simulation and not on performance data of real plants. Therefore, validation of FC performance against the reference curves utilised would be required to determine the effects on the simulated overall system performance.

The assumptions, scale and assessment methodologies varied from study to study so an attempt has been made here to normalise the results with key differences in performance assumptions highlighted.

5.3. Techno-Economic Data Rationalisation

The reference base cases and cases identified in the literature review were conducted with different currency types, plant locations and currency year. To ensure a like for like comparison the reported capital costs were scaled to 2017 Euros for a plant in the Netherlands.

The scaling factors used for rationalisation considered plant location factors for Material, Labour (Productivity), Labour (Cost) and Contingency (IEAGHG, 2018) as shown in Table 6. The equipment cost factor was taken from an in-house database, currency conversion was taken as the yearly average exchange rate from (OFX, 2018) and a currency inflation factor was taken from annual International Monetary Fund Consumer Price Index data (IMF, 2018).

Table 6: Plant Location Factors (IEAGHG, 2018)

Location	Material Factors	Labour Factors (productivity)	Labour Factors (costs)	Contingency
The Netherlands	1.00	1.00	1.00	10%
Eastern Europe	0.92	1.28	0.40	10%
USA (mid-west)	0.94	0.95	0.92	10%
Canada	1.07	1.12	0.96	10%
South Africa	1.03	2.24	0.70	10%
Australia	1.00	1.23	1.38	10%
India	0.93	2.42	0.26	15%
China	0.77	2.29	0.16	15%
Japan	0.91	0.98	0.68	10%
South-East Asia	0.92	1.78	0.24	15%
South America	0.97	1.54	0.28	20%
Middle East	0.91	1.84	0.24	20%

The scaling factors outlined in Table 6 were applied to the capital costs based on the breakdown of TPC as presented in Section 5.2 for the given plant / site location. For example the material factor was only applied to the materials aspect of the capital cost (10% of BEC), Labour productivity and cost factors were applied to the labour aspect of the capital cost (30% BEC) etc. The currency conversion factor was applied for all the capital cost and the currency inflation factor was applied to all the capital cost elements except the equipment cost as this factor takes into account the escalation or de-escalation of costs of various plant items relative to the two years considered.

5.4. Power Plant Output Scaling

All cases here have been rationalised to ensure they have the same net power output of 634 MW_e, for the purposes of comparison no account has been made of the potential FC modular sizes or other power generating equipment standard sizes. The fuel cells and other generating plant have been assumed to produce the exact amount of power required to deliver the 634 MWe net at the efficiencies stated in the respective papers. This enables all cases to be judged on a consistent basis. In addition the comparison has assumed that both SOFC and MCFC technology are available at the scales required despite the potential scale up issues, particularly with regards to SOFC, as highlighted in Section 2.1.

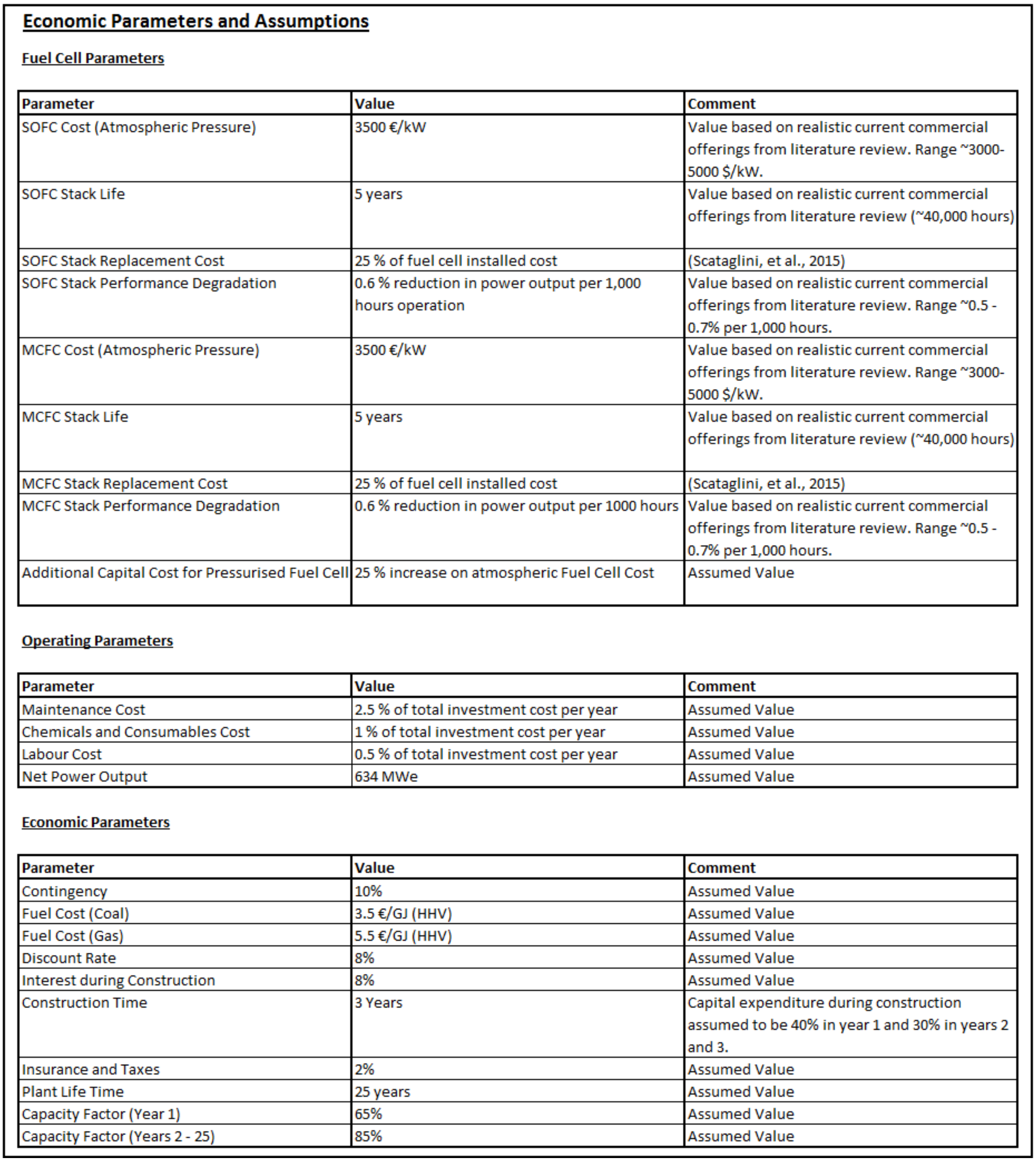
The data for the cases identified in the literature review and reference base cases were presented for a range of net power plant outputs between ~250 and 935 MW_e. Given that the plants are all of the same order of magnitude the rationalised costs have been scaled linearly on a €/kW_e basis. The reported net plant efficiency on a HHV basis and the CO_2 emissions in terms of g/kWh of the scaled plant has been assumed to be the same as the plant prior to scaling for all cases.

The breakdown of the plant costs for all of the cases considered in the techno-economic analysis, both before and after rationalisation, are summarised in Appendix 3 as presented in the following tables:-

- Table A3-2:- Summary of Installed Costs Pre-Rationalisation
- Table A3-3:- Summary of Installed Costs Post-Rationalisation
- Table A3-4:- Summary of Installed Costs per kW Post-Rationalisation
- Table A3-5:- Summary of Installed Costs Post-Rationalisation and Scaling

5.5. Techno-Economic Assessment Parameters

A set of common parameters were used for all cases considered in the techno-economic assessment they are detailed in Figure 46.



Economic Parameters and Assumptions

Fuel Cell Parameters

Parameter	Value	Comment
SOFC Cost (Atmospheric Pressure)	3500 €/kW	Value based on realistic current commercial offerings from literature review. Range ~3000-5000 $/kW.
SOFC Stack Life	5 years	Value based on realistic current commercial offerings from literature review (~40,000 hours)
SOFC Stack Replacement Cost	25 % of fuel cell installed cost	(Scataglini, et al., 2015)
SOFC Stack Performance Degradation	0.6 % reduction in power output per 1,000 hours operation	Value based on realistic current commercial offerings from literature review. Range ~0.5 - 0.7% per 1,000 hours.
MCFC Cost (Atmospheric Pressure)	3500 €/kW	Value based on realistic current commercial offerings from literature review. Range ~3000-5000 $/kW.
MCFC Stack Life	5 years	Value based on realistic current commercial offerings from literature review (~40,000 hours)
MCFC Stack Replacement Cost	25 % of fuel cell installed cost	(Scataglini, et al., 2015)
MCFC Stack Performance Degradation	0.6 % reduction in power output per 1000 hours	Value based on realistic current commercial offerings from literature review. Range ~0.5 - 0.7% per 1,000 hours.
Additional Capital Cost for Pressurised Fuel Cell	25 % increase on atmospheric Fuel Cell Cost	Assumed Value

Operating Parameters

Parameter	Value	Comment
Maintenance Cost	2.5 % of total investment cost per year	Assumed Value
Chemicals and Consumables Cost	1 % of total investment cost per year	Assumed Value
Labour Cost	0.5 % of total investment cost per year	Assumed Value
Net Power Output	634 MWe	Assumed Value

Economic Parameters

Parameter	Value	Comment
Contingency	10%	Assumed Value
Fuel Cost (Coal)	3.5 €/GJ (HHV)	Assumed Value
Fuel Cost (Gas)	5.5 €/GJ (HHV)	Assumed Value
Discount Rate	8%	Assumed Value
Interest during Construction	8%	Assumed Value
Construction Time	3 Years	Capital expenditure during construction assumed to be 40% in year 1 and 30% in years 2 and 3.
Insurance and Taxes	2%	Assumed Value
Plant Life Time	25 years	Assumed Value
Capacity Factor (Year 1)	65%	Assumed Value
Capacity Factor (Years 2 - 25)	85%	Assumed Value

Figure 46: Techno-Economic Analysis Economic Parameters and Assumptions

5.6. Techno-Economic Assessment

For each case considered the levelised cost of electricity (LCOE) is calculated by modelling the net present value (NPV) of the plant's cash flow and adjusting the assumed electricity price stated in the model until it gives a zero NPV utilising equation 5-1.

$$NPV = \frac{FV_0}{(1+r_0)^{t_0}} + \frac{FV_1}{(1+r_1)^{t_1}} + \frac{FV_2}{(1+r_2)^{t_2}} + \cdots + \cdots + \frac{FV_n}{(1+r_n)^{t_n}} \tag{5-1}$$

where,

FV (Future Value) = net cash inflow-outflows expected during a particular period

R = discount rate or return that could be earned in alternative investments

t = time period

The data for each case was rationalised using the same method and cash flow modelled based on the economic parameters presented in Figure 46. The primary value of this approach lies not in the absolute accuracy of cost estimates for the individual cases, but in the fact that all cases were evaluated using a common methodology with a consistent set of economic assumptions to ensure a fair comparison can be made between the various technologies. The calculated LCOE is presented in € cent/kWh.

In addition to LCOE, the Cost of CO_2 avoided is calculated which represents the minimum emissions price that, when applied to both a power plant with CO_2 capture and without CO_2 capture, would incentivise the power plant with CO_2 capture. The cost is reported in €/tonne and is calculated as follows:

$$Cost\ of\ CO_2\ Avoided = \frac{(\ LCOE_{\ \mathrm{CCS}} - LCOE_{\ \mathrm{Non\text{-}CCS}})}{(\ CO_2\ Emissions_{\ \mathrm{Non\text{-}CCS}} - CO_2\ Emissions_{\ \mathrm{CCS}})} \qquad (5\text{-}2)$$

Where:

- CCS is the CO_2 Capture plant Case for which the cost of CO_2 avoided is calculated
- Non-CCS is the Reference Case with no CO_2 capture plant
- LCOE is the levelised cost of Electricity in €/MWh
- CO_2 Emissions are the rate of CO_2 emitted from the stack in tonne/MWh

A limitation with this study's approach is that base capital cost data was taken from a number of different studies collected in the literature review.

The capital cost data has been homogenised in terms of location, currency, year etc. However the raw data taken from the literature review was unchanged. Therefore a degree of variability is introduced from the different studies' base data and the accuracy and robustness of capital cost estimates they employ. This could potentially lead to varying results with comparatively optimistic or pessimistic cost assumptions. It should be noted that for all cases the base data for the FC capital cost was adjusted based on the findings of the literature review.

The results presented here, referenced to common base cases, are intended to provide a guide towards the most promising configurations and technologies within the described limits. For completeness the cost of CO_2 avoided has been included for each case as reported in the individual studies, it should be noted that these costs have not been adjusted or homogenised.

It is recommended that the cases are modelled to confirm performance as per literature with bottom up estimates made for each case using a single source of data. This will ensure that the results are developed using the same basis and methodology and are therefore directly comparable. However this was beyond the scope of this present study. The costs associated with CO_2 transport and storage have been excluded. There is a significant variation in costs depending on the storage option. The 2011 € storage cost was in the range 1 to 7 €/tonne CO_2 for onshore depleted oil and gas fields to 6 to 20 €/tonne CO_2 for offshore deep saline aquifers (IEAGHG, 2011). The 2016 transport and storage cost was estimated to be in the range 8 to 31 £/tonne CO_2 (Leigh Fisher & Jacobs, 2016).

For the purposes of this study, no value has been given for the cost of emitting CO_2 (CO_2 Tax) or value of CO_2 captured (emissions trading scheme credit / CO_2 utilisation).

5.7. Reference Case 1a Assessment: SC PC without CO_2 Capture

5.7.1. Plant Description

Reference Case 1a considers SC PC boiler plant with no form of carbon capture technology. Coal is fed to a supercritical boiler with single re-heat cycle that raises steam to 24.1MPa/593°C/593°C.

The overall SC PC plant configuration is shown in Figure 47 (NETL, 2015).

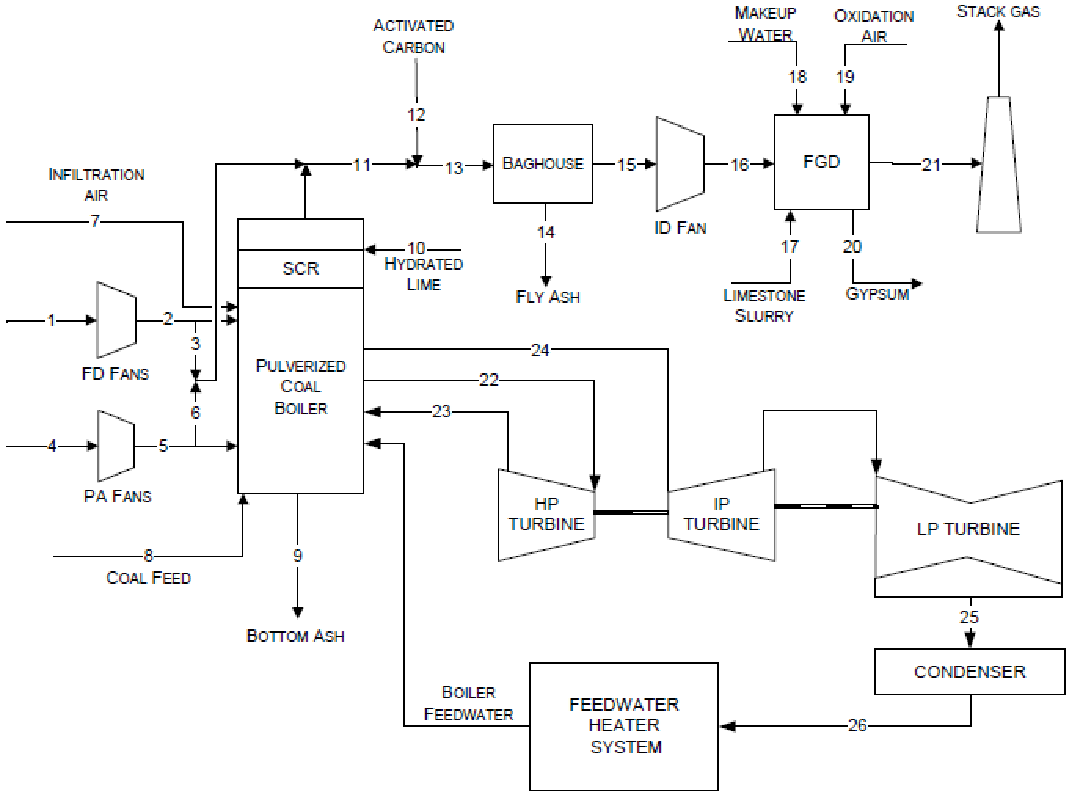


Figure 47: Reference Case 1a: Supercritical PC Boiler Plant Configuration (NETL, 2015)

5.7.2. Plant Performance

Reference Case 1a produces a net output of 634 MW_e at a net plant efficiency of 40.7 percent (HHV basis), with plant CO_2 emissions of 774 g/kWh. This forms the base non-CCS reference case for options that consider PC technology for the calculation of the cost of CO_2 avoidance.

5.7.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-1. For Reference Case 1a, LCOE was calculated at 9.61 € cent/kWh. As there is no CO_2 capture, the cost of CO_2 avoided is not applicable.

5.8. Reference Case 1b Assessment: SC PC with CO_2 Capture

5.8.1. Plant Description

Reference Case 1b considers the Supercritical PC boiler plant from Reference Case 1a with the addition of CCS, utilising Cansolv's CO_2 Capture Process.

Cansolv's CO_2 capture process consists of a single, rectangular concrete absorber containing stainless-steel packing. Flue gas enters the absorber and flows counter current to the Cansolv amine-based solvent and CO_2 is absorbed into the lean solvent. The CO_2 rich solvent is then sent to the stainless steel stripper vessel which contains structured stainless steel packing. The CO_2 rich amine is heated using low pressure steam to releases the CO_2 product. The lean amine is then returned to the absorber. A typical flow sheet for the Cansolv process is shown in Figure 48 (NETL, 2015).

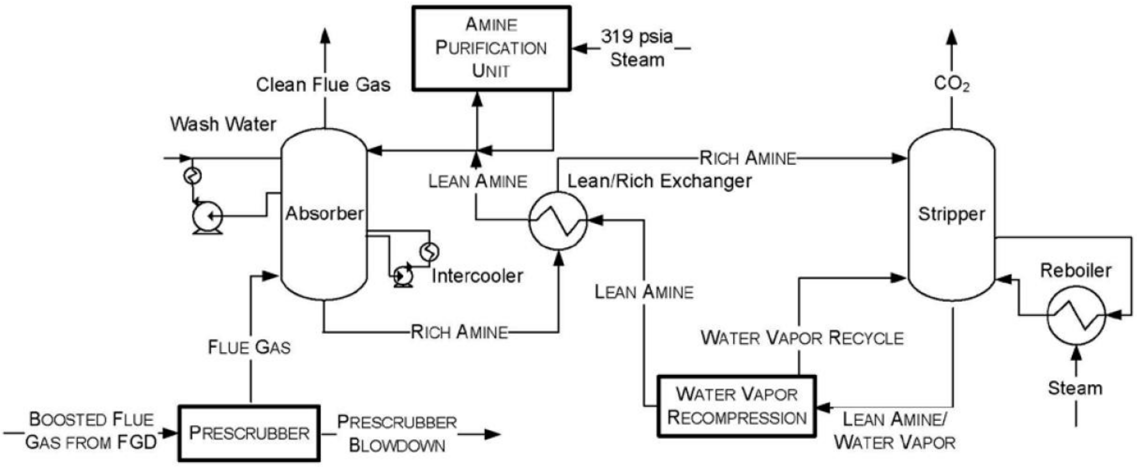


Figure 48: Cansolv Amine Scrubbing Plant Configuration (NETL, 2015)

The overall SC PC with CO_2 Capture plant configuration is shown in Figure 49 (NETL, 2015).

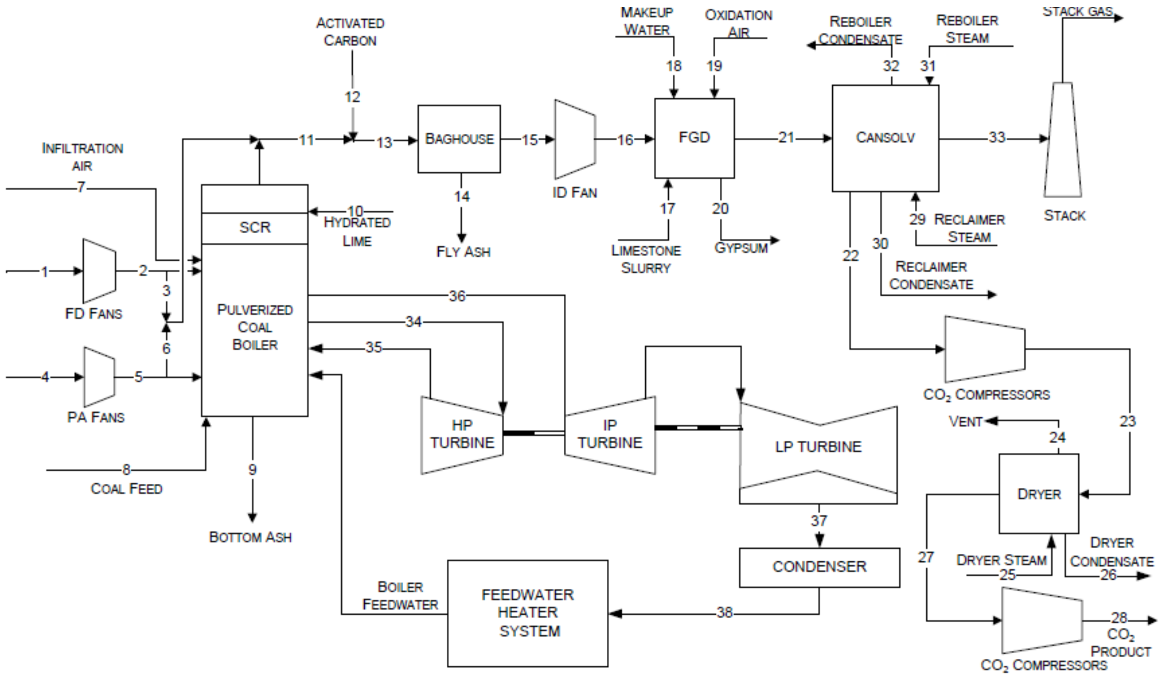


Figure 49: Reference Case 1b: Supercritical PC Boiler Plant plus Cansolv Amine Scrubbing Plant Configuration (NETL, 2015)

5.8.2. Plant Performance

Reference Case 1b produces a net output of 634 MW_e at a net plant efficiency of 32.5 percent (HHV basis). 90% of the CO_2 in the flue gas is removed by the CO_2 capture plant, with plant CO_2 emissions of 97 g/kWh. This forms the base CCS technology reference case for options that consider utilising PC boiler plant.

5.8.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-2. For Reference Case 1b, LCOE was calculated at 15.20 € cent/kWh with the cost of CO_2 avoided 82.6 €/t CO_2 relative to SC PC Reference Case 1a.

5.9. Reference Case 2a Assessment: NGCC without CO_2 Capture

5.9.1. Plant Description

Reference Case 2a considers a NGCC power plant with no form of carbon capture technology. The design considers two 2013 F- Class combustion turbine generators, two heat recovery steam generators (HRSGs) and one steam turbine. The single reheat steam cycle raises steam to 16.5MPa/566°C/566°C.

The overall NGCC without CO_2 Capture plant configuration is shown in Figure 50 (NETL, 2015).

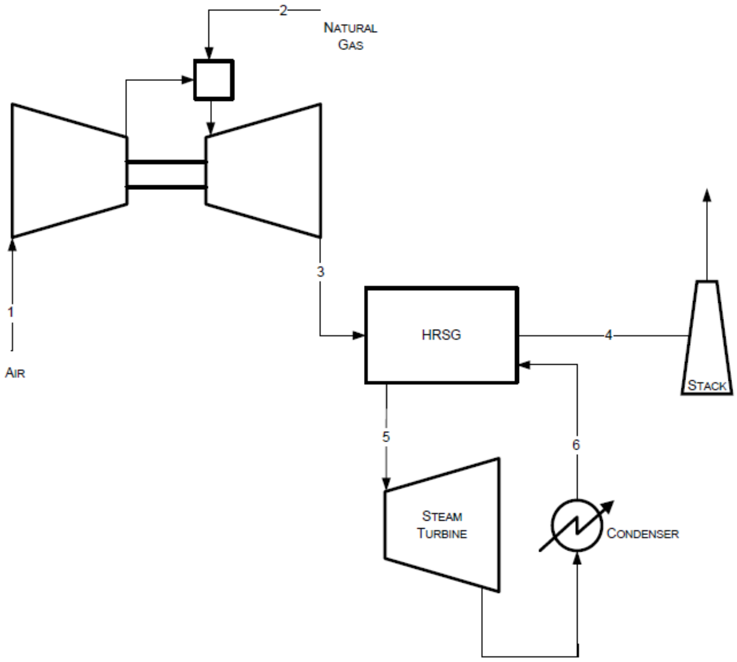


Figure 50: Reference Case 2a: NGCC Power Plant Configuration (NETL, 2015)

5.9.2. Plant Performance

Reference Case 2a produces a net output of 634 MW_e at a net plant efficiency of 51.5 percent (HHV basis), with plant CO_2 emissions of 357 g/kWh. This forms the base non-CCS reference case for options that consider NGCC technology for the calculation of the cost of CO_2 avoidance.

5.9.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-3. For Reference Case 2a, LCOE was calculated at 6.05 € cent/kWh. As there is no CO_2 capture, the cost of CO_2 avoided is not applicable.

5.10. Reference Case 2b Assessment: NGCC with CO_2 Capture

5.10.1. Plant Description

Reference Case 2b considers the NGCC plant from Reference Case 2a with the addition of CCS, utilising Cansolv's CO_2 Capture Process as described in Reference Case 1b.

The overall NGCC with CO_2 Capture plant configuration is shown in Figure 51 (NETL, 2015).

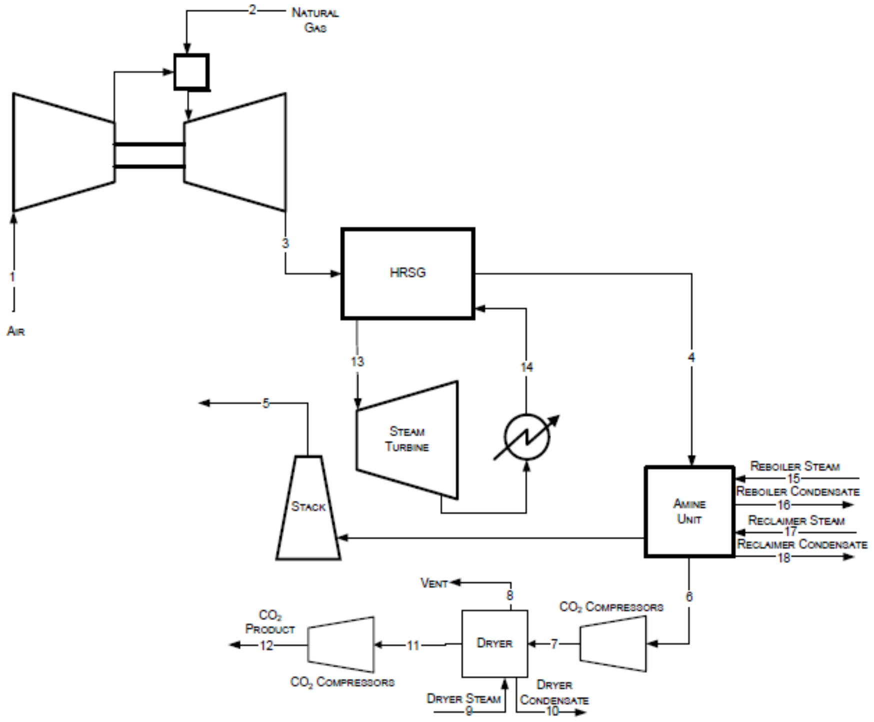


Figure 51: Reference Case 2b: NGCC Power Plant plus Cansolv Amine Scrubbing Plant Configuration (NETL, 2015)

5.10.2. Plant Performance

Reference Case 2b produces a net output of 634 MW_e at a net plant efficiency of 45.7 percent (HHV basis). 90% of the CO_2 in the flue gas is removed by the CO_2 capture plant, with plant CO_2 emissions of 40 g/kWh. This forms the base CCS technology reference case for options that consider utilising a NGCC plant.

5.10.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-4. For Reference Case 2b, LCOE was calculated at 9.09 € cent/kWh with the cost of CO_2 avoided 96.0 €/t CO_2 relative to NGCC Reference Case 2a.

5.11. Reference Case 3a Assessment: IGCC without CO_2 Capture

5.11.1. Plant Description

Reference Case 3a considers an IGCC power plant with no form of carbon capture technology. The design considers General Electric Energy (GEE) gasifier technology fed with coal slurry and oxygen produced in an air separation unit (ASU) to produce the syngas. Syngas is then cooled and cleaned in a number of stages, before being fed to a 2008 F-Class combustion turbine, a HRSG recovers heat from the flue gas and generates steam for the steam turbine. The single reheat steam cycle raises steam to 12.4MPa/562°C/562°C.

The overall IGCC without CO_2 Capture plant configuration is shown in Figure 52 (NETL, 2015).

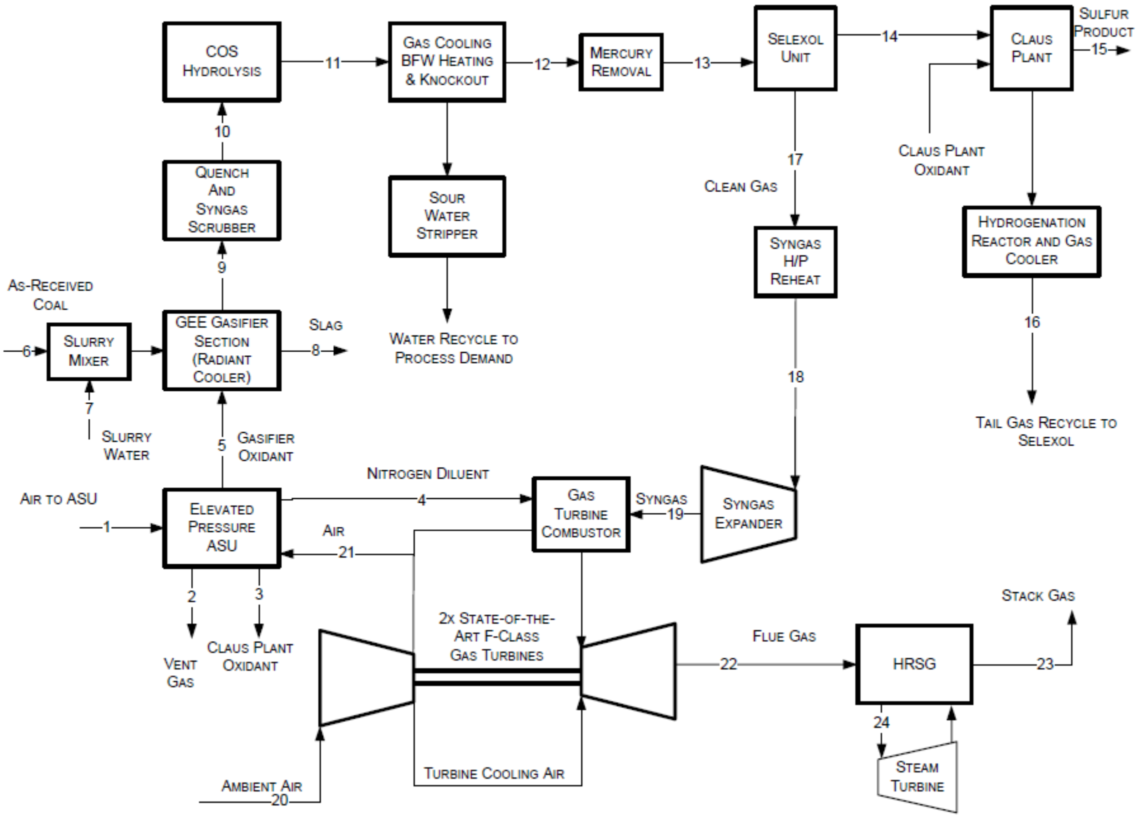


Figure 52: Reference Case 3a: IGCC Power Plant Configuration (NETL, 2015)

5.11.2. Plant Performance

Reference Case 3a produces a net output of 634 MW_e at a net plant efficiency of 39 percent (HHV basis), with plant CO_2 emissions of 782 g/kWh. This forms the base non-CCS reference case for options that consider IGCC technology for the calculation of the cost of CO_2 avoidance.

5.11.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-5. For Reference Case 3a, LCOE was calculated at 11.01 € cent/kWh. As there is no CO_2 capture, the cost of CO_2 avoided is not applicable.

5.12. Reference Case 3b Assessment: IGCC with CO_2 Capture

5.12.1. Plant Description

Reference Case 3b considers the IGCC plant from Reference Case 3a with the addition of CCS, utilising a two-stage selexol process.

The syngas enters the first absorber in which H_2S is removed using some of the loaded solvent from the CO_2 absorber. The gas exiting the H_2S absorber passes to the second absorber where CO_2 is removed using flash regenerated, chilled solvent and thermally regenerated solvent. The CO_2 loaded solvent leaves the CO_2 absorber and is regenerated through a series of flash drums. The CO_2 is captured as a product stream from the flash drums. A typical flow sheet for the two-stage selexol process is shown in Figure 53 (NETL, 2015).

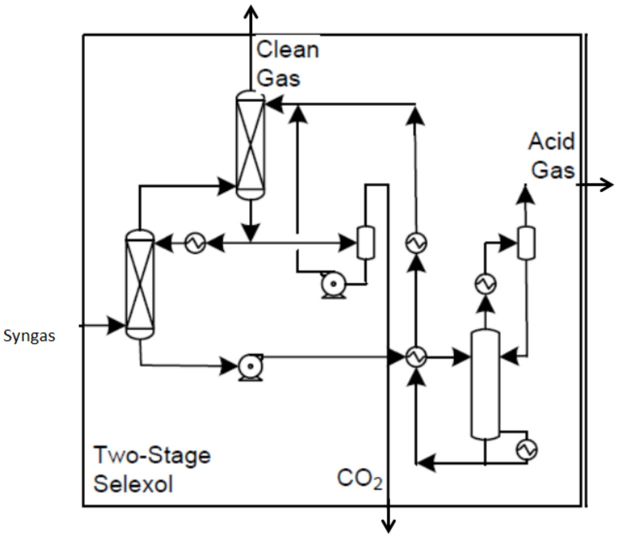


Figure 53: Two-Stage Selexol Plant Configuration (NETL, 2015)

5.12.2. Plant Performance

Reference Case 3b produces a net output of 634 MW_e at a net plant efficiency of 32.6 percent (HHV basis). 90% of the CO_2 in the flue gas is removed by the CO_2 capture plant, with plant CO_2 emissions of 93 g/kWh. This forms the base CCS technology reference case for options that consider utilising an IGCC plant.

5.12.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-6. For Reference Case 3b, LCOE was calculated at 14.74 € cent/kWh with the cost of CO_2 avoided 54.2 €/t CO_2 relative to Reference Case 3a.

5.13. Case 1 Assessment: Atmospheric NGFC-MCFC with CO_2 Capture

5.13.1. Plant Description

Case 1 considers a hybrid cycle consisting of a NGCC power plant and an atmospheric pressure MCFC equipped with cryogenic carbon capture technology. The design considers a MCFC placed between the gas turbine and HRSG.

Natural gas undergoes desulphurisation in filter beds, using active carbon with metal impregnation, and is fed to the MCFC anode where the natural gas is internally reformed. The MCFC cathode is fed with the gas turbine exhaust gases and CO_2 is transferred to the anode side which concentrates the CO_2 in the anode effluent.

To capture the CO_2 a cryogenic process is utilised, in which the anode effluent stream is cooled down to a temperature approaching the triple point of CO_2 (-56.6°C). Most of the CO_2 condenses and can be separated by gravity; other components with a much lower boiling point remain in the gas phase.

The overall Atmospheric NGFC-MCFC with CO_2 Capture plant configuration is shown in Figure 54 (Campanari, et al., 2014).

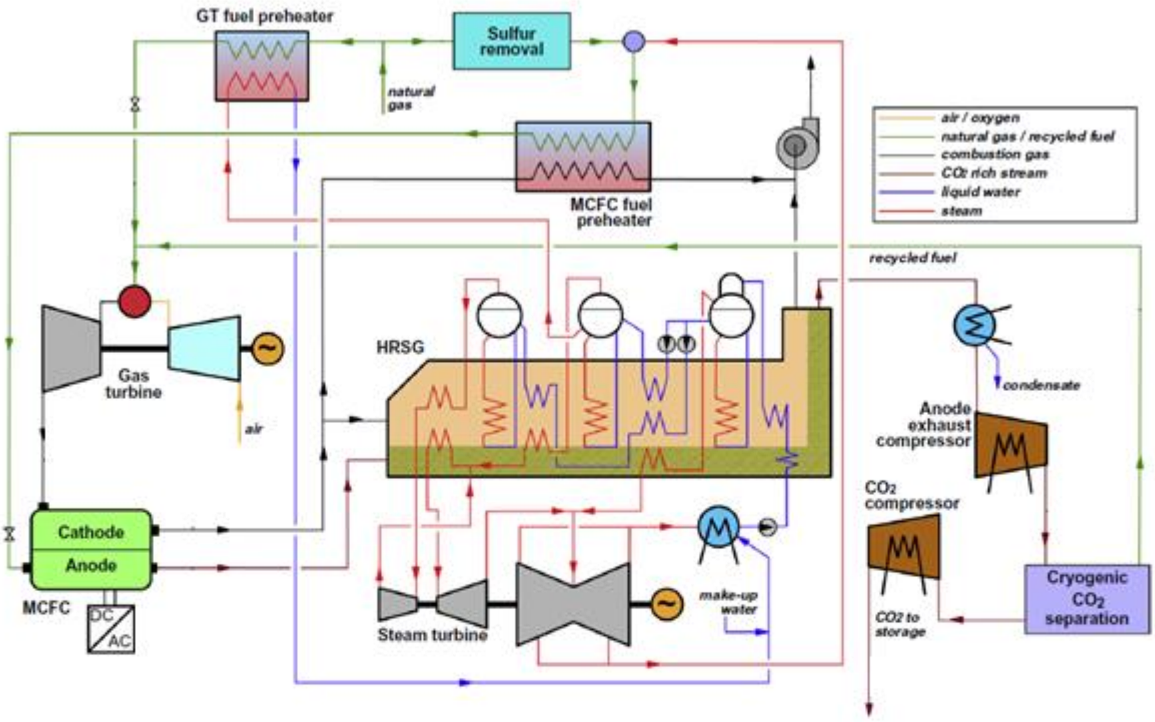


Figure 54: Case 1: NGCC plus MCFC Hybrid-Cycle Configuration (Campanari, et al., 2014)

5.13.2. Plant Performance

The FC operational parameters are shown in Table 7.

Table 7: Case 1 FC Operational Parameters

Parameter	Value
Voltage	0.709 V
U_f	75%
U_{CO2}	75%
Temperature	650°C
Pressure	1 bara
MCFC Capacity	119 MW_e

Case 1 produces a net output of 634 MW_e at a net plant efficiency of 58.9 percent (HHV basis), 71.9% of CO_2 is captured with plant CO_2 emissions of 98 g/kWh.

5.13.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-7. For Case 1, LCOE was calculated at 6.92 € cent/kWh with the cost of CO_2 avoided 33.5 €/t CO_2 relative to NGCC Reference Case 2a. For comparison the unadjusted cost of CO_2 avoided, as determined from the results reported in (Campanari, et al., 2014) was 120.4 €/t CO_2.

5.14. Case 2 Assessment: Atmospheric IGFC-SOFC with CO_2 Capture

5.14.1. Plant Description

Case 2 considers a hybrid cycle consisting of a catalytic coal gasifier, heat recovery and dry gas cleaning. The syngas is then fed to an atmospheric pressure SOFC equipped with carbon capture technology in the form of oxy-combustion, heat recovery and condensation of the anode off gas.

Syngas produced in the catalytic gasifier undergoes dry cleaning process to remove solids, halides, sulphur and mercury prior to being fed to the SOFC anode where the syngas is internally reformed. The SOFC cathode is fed with air.

To capture the CO_2 an oxy-combustion process is utilised, in which the anode effluent stream is combusted in the presence of 95% pure O_2 resulting in a stream of CO_2, O_2 and H_2O. The stream is then purified through condensation.

The overall Atmospheric IGFC-SOFC with CO_2 Capture plant configuration is shown in Figure 55 (NETL, 2009).

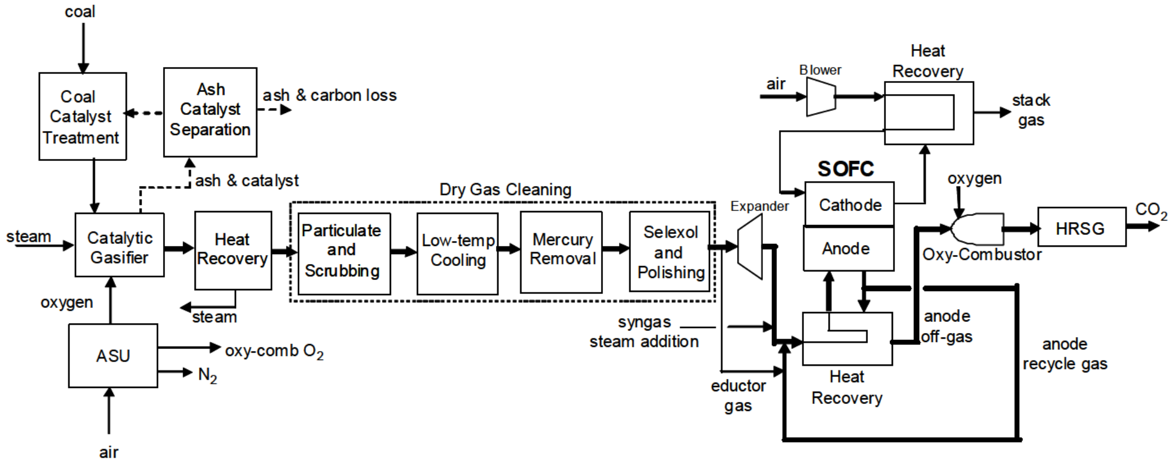


Figure 55: Case 2: IGCC plus atmospheric SOFC Hybrid-Cycle Configuration (NETL, 2009)

5.14.2. Plant Performance

The FC operational parameters are shown in Table 8.

Table 8: Case 2 FC Operational Parameters

Parameter	Value
Voltage	0.8 V
U_f	82%
Temperature	650°C
Pressure	1.8 bara
SOFC Capacity	598 MW_e

Case 2 produces a net output of 634 MW_e at a net plant efficiency of 49.4 percent (HHV basis), >99% of CO_2 is captured with plant CO_2 emissions of 1 g/kWh.

5.14.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-8. For Case 2, LCOE was calculated at 19.18 € cent/kWh with the cost of CO_2 avoided 104.7 €/t CO_2 relative to IGCC Reference Case 3a. For comparison the unadjusted cost of CO_2 avoided, as determined from the results reported in (NETL, 2009) was -31.3 US$/t CO_2.

5.15. Case 3 Assessment: Pressurised IGFC-SOFC with CO_2 Capture

5.15.1. Plant Description

Case 3 considers a hybrid cycle consisting of a radiant only coal gasifier, heat recovery, syngas shifting and syngas cleaning. The syngas is then fed to a pressurised SOFC equipped with carbon capture technology in the form of adiabatic oxidation, heat recovery and condensation of the anode off gas.

Syngas produced in the gasifier undergoes a water gas shift reaction to convert CO in the syngas which has a detrimental effect on SOFC performance. The syngas then undergoes cleaning to remove solids, halides, sulphur and mercury prior to being fed to the SOFC anode. The SOFC cathode is fed with air.

To capture the CO_2 the anode effluent stream is fed to an adiabatic oxidation unit in which it is reacted with a stoichiometric amount of O_2 from the ASU to produce a stream primarily containing a mixture of CO_2 and H_2O. The stream is then purified through condensation via a series of flash cascade drums.

The overall Pressurised IGFC-SOFC with CO_2 Capture plant configuration is shown in Figure 56 (Adams & Barton, 2010)

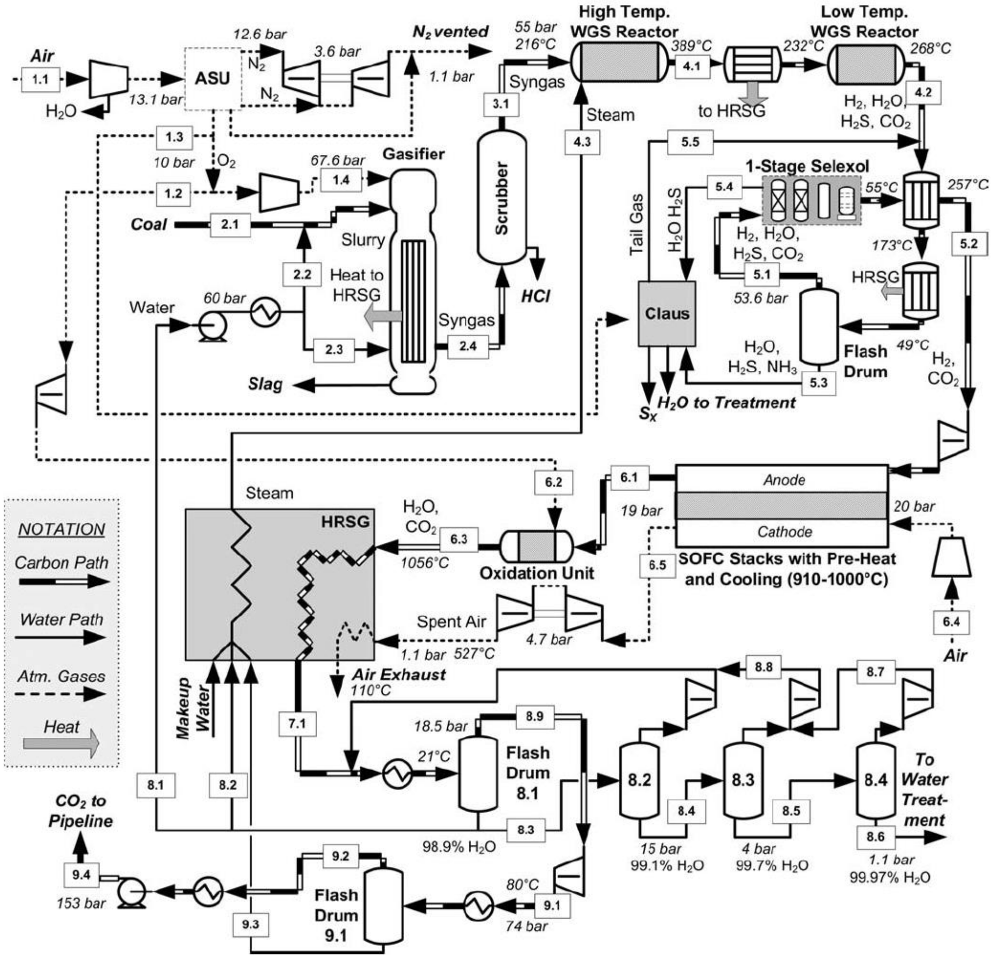


Figure 56: Case 3: IGCC plus pressurised SOFC Hybrid-Cycle Configuration (Adams & Barton, 2010)

5.15.2. Plant Performance

The FC operational parameters are shown in Table 9.

Table 9: Case 3 FC Operational Parameters

Parameter	Value
Voltage	0.69 V
U_f	99.5%
Temperature	1000°C
Pressure	20 bara
SOFC Capacity	466 MW_e

Case 3 produces a net output of 634 MW_e at a net plant efficiency of 44.8 percent (HHV basis), 100% of CO_2 is captured with plant CO_2 emissions of 0 g/kWh.

5.15.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-9. For Case 3, LCOE was calculated at 18.75 € cent/kWh with the cost of CO_2 avoided 99.0 €/t CO_2 relative to IGCC Reference Case 3a. For comparison the unadjusted cost of CO_2 avoided, as determined from the results reported in (Adams & Barton, 2010) was 6.1 US$/t CO_2.

5.16. Case 4 Assessment: Pressurised NGFC-SOFC with CO_2 Capture

5.16.1. Plant Description

Case 4 considers a hybrid cycle consisting of hydrogen produced from natural gas in steam-methane reformers fed to a pressurised SOFC equipped with carbon capture technology in the form of adiabatic oxidation, heat recovery and condensation of the anode off gas.

Natural gas is reformed with steam to produce hydrogen, this gas then undergoes a water gas shift reaction to convert CO in the gas which has a detrimental effect on SOFC performance. The gas is then fed to the SOFC anode. The SOFC cathode is fed with air.

To capture the CO_2 the anode effluent stream is fed to an adiabatic oxidation unit in which it is reacted with a stoichiometric amount of O_2 from an ASU to produce a stream primarily containing a mixture of CO_2 and H_2O. The stream is then purified through condensation via a series of flash cascade drums.

The overall Pressurised NGFC-SOFC with CO_2 Capture plant configuration is shown in Figure 57 (Adams & Barton, 2010)

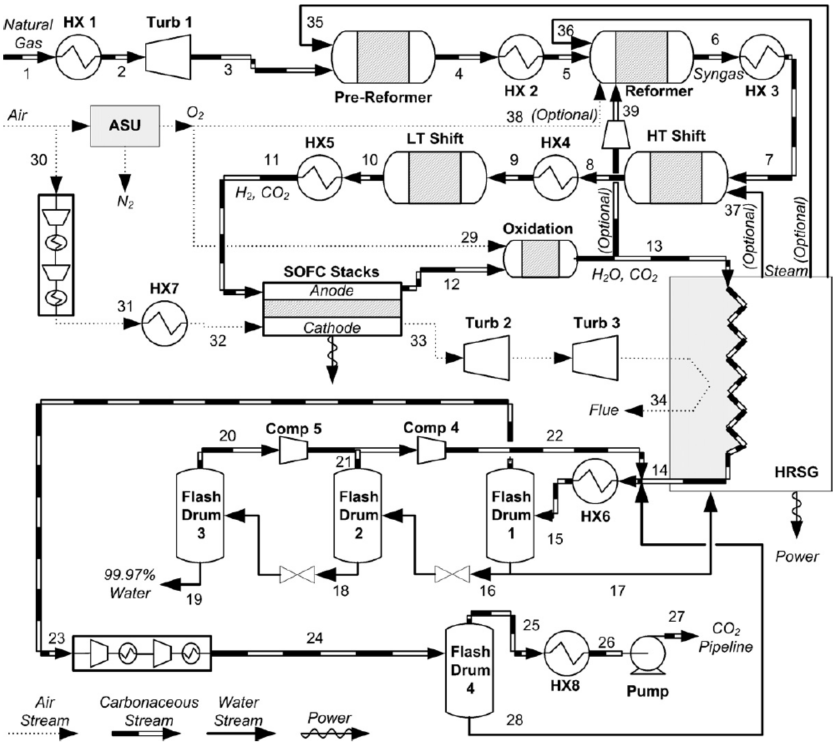


Figure 57: Case 4: NG plus SOFC Hybrid-Cycle Configuration (Adams & Barton, 2010)

5.16.2. Plant Performance

The FC operational parameters are shown in Table 10.

Table 10: Case 4 FC Operational Parameters

Parameter	Value
Voltage	0.69 V
U_f	92%
Temperature	950°C
Pressure	10.1 bara
SOFC Capacity	688 MW_e

Case 4 produces a net output of 634 MW_e at a net plant efficiency of 74 percent (HHV basis), 100% of CO_2 is captured with plant CO_2 emissions of 0 g/kWh.

5.16.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-10. For Case 4, LCOE was calculated at 19.55 € cent/kWh with the cost of CO_2 avoided 378.2 €/t CO_2 relative to NGCC Reference Case 2a. For comparison the unadjusted cost of CO_2 avoided, as determined from the results reported in (Adams & Barton, 2010) was -6.1 US$/t CO_2.

5.17. Case 5 Assessment: Atmospheric Hybrid-MCFC with CO_2 Capture

5.17.1. Plant Description

Case 5 considers a hybrid cycle consisting of consisting of a coal fired power plant and an atmospheric pressure MCFC equipped with carbon capture technology in the form of oxy-combustion and condensation.

Coal is fired in a SC PC boiler with air, and following removal of ash, SOx and NOx, the exhaust gas is fed to the MCFC cathode. Natural gas is reformed externally before being fed to the MCFC anode. CO_2 is transferred from the cathode side of the MCFC to the anode side which concentrates the CO_2 in the anode effluent.

To capture CO_2, an oxy-combustion process is utilised, in which the anode effluent stream is combusted in the presence of 95% pure O_2 resulting in a stream of predominantly CO_2 and H_2O. The stream is then purified through condensation.

The overall Atmospheric Hybrid-MCFC with CO_2 Capture plant configuration is shown in Figure 58 (Duan, et al., 2016).

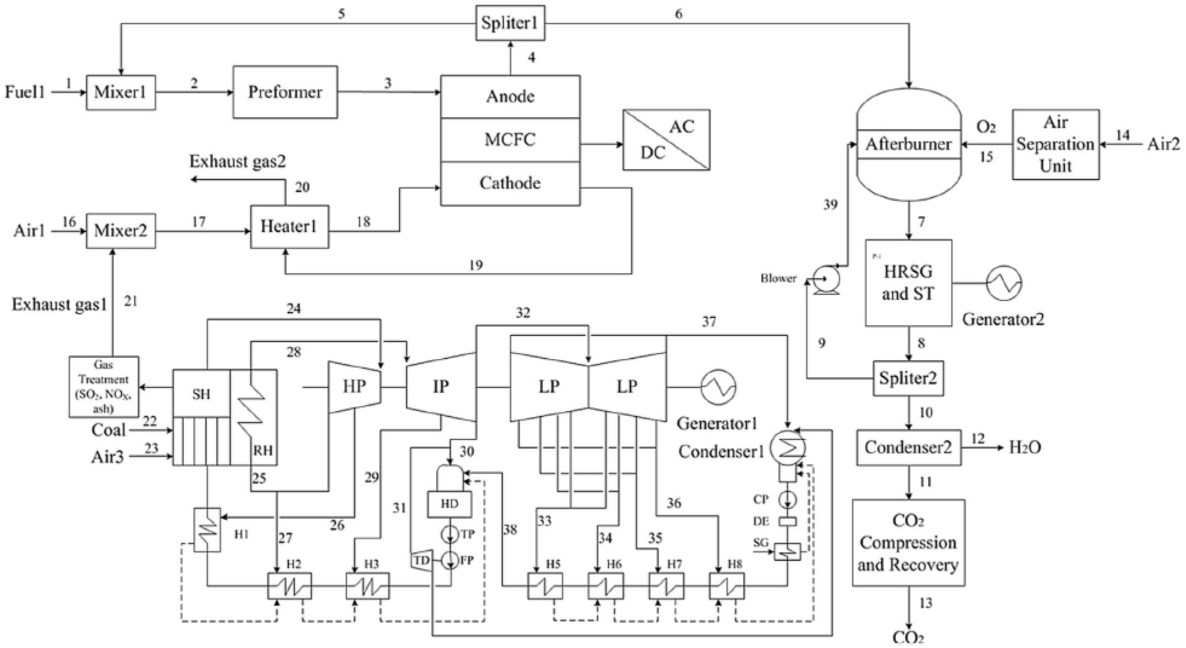


Figure 58: Case 5: SC PC Power Plant plus MCFC Hybrid-Cycle Configuration (Duan, et al., 2016)

5.17.2. Plant Performance

The FC operational parameters are shown in Table 11.

Table 11: Case 5 FC Operational Parameters

Parameter	Value
Voltage	0.68 V
U_f	85%
U_{CO2}	52%
Temperature	650°C
Pressure	1 bara
MCFC Capacity	213 MWe

Case 5 produces a net output of 634 MW_e at a net plant efficiency of 45.6 percent (HHV basis), 88% of CO_2 is captured with plant CO_2 emissions of 113 g/kWh.

5.17.3. Economic Assessment

The economic assessment was carried out as described above and the results are shown in Appendix 3 Figure A3-11. For Case 5, LCOE was calculated at 8.62 € cent/kWh with the cost of CO_2 avoided -15.0 €/t CO_2 relative to SC PC Reference Case 1a. For comparison the unadjusted cost of CO_2 avoided, as determined from the results reported in (Duan, et al., 2016) was 60.7 US$/t CO_2.

5.18. Case Performance Summary

The results of each case are summarised in Table 12.

Table 12: Summary of Key Techno-Economic Analysis Results: Reference Cases and Cases 1-5

Case	Ref 1a	Ref 1b	Ref 2a	Ref 2b	Ref 3a	Ref 3b	1	2	3	4	5
Description	SC PC w/o CCS	SC PC with CCS	NGCC w/o CCS	NGCC with CCS	IGCC w/o CCS	IGCC with CCS	Atmospheric NG-MCFC with CCS	Atmospheric IG-SOFC with CCS	Pressurised IG-SOFC with CCS	Pressurised NG-SOFC with CCS	Atmospheric SC PC MCFC with CCS
Performance											
Net Power Output (MWe)	634	634	634	634	634	634	634	634	634	634	634
Net Plant HHV efficiency (%)	40.7	32.5	51.5	45.7	39.0	32.6	58.9	49.4	44.8	74.0	45.6
HHV Thermal Input (MWth)	1557.7	1950.8	1231.1	1387.3	1625.6	1944.8	1076.8	1283.4	1415.2	856.8	1389.7
CO_2 emissions (g/kWh)	774	97	357	40	782	93	98	1	0	0	113
CO_2 Capture (%)	0	90	0	90	0	90	72	>99	100	100	88
Cost											
Installed cost (2017 M€)	1653.4	2875.1	558.6	1208.7	1974.2	2761.3	800.7	3164.7	3234.2	3367.3	1185.1
LCOE (2017 € cent/kWh)	9.61	15.20	6.05	9.09	11.01	14.74	6.92	19.18	18.75	19.55	8.62
Cost of CO_2 avoided (2017 € /t CO_2)	N/A	82.6	N/A	96.0	N/A	54.2	33.5	104.7	99.0	378.2	-15.0
Ref Case for Cost of CO_2 Avoided	N/A	Ref 1a	N/A	Ref 2a	N/A	Ref 3a	Ref 2a	Ref 3a	Ref 3a	Ref 2a	Ref 1a

Case 1 (Atmospheric NGFC-MCFC with CO_2 Capture) gave the best performance of all the cases with CO_2 capture with a LCOE of 6.92 € cent/kWh (2017 €), a 23.9% decrease on the equivalent NGCC plus Cansolv PCC Case 2b. This shows that MCFCs as a method of CO_2 capture from a NGCC have the potential to be significantly more cost effective than the current state-of-the-art proprietary solvent technologies.

Case 5 (Atmospheric Hybrid-MCFC with CO_2 Capture) shows a LCOE performance of a similar order of magnitude as NGCC Reference Case 2b, albeit with a much better performance in terms of cost of CO_2 avoided. This cost of CO_2 avoided result is referred to the SC PC Reference Case 1a (as Case 5 primarily considers SC PC boiler technology). When referenced to the reference base case with the lowest starting CO_2 emissions (and LCOE), NGCC Case 2a, the revised cost of CO_2 avoided is 105.1 €/t CO_2 which is more in line with the other CO_2 capture cases. The negative cost of CO_2 avoided derived here is likely caused by the use of data from two different studies with either optimistic or pessimistic cost assumptions used. When the unadjusted reported costs of CO_2 avoided from the individual studies are considered, there is a large variation in the results from those presented here. One potential reason for these discrepancies is that, for each case, the FC capital cost was adjusted to match the findings from the literature review. The FC capital cost is a significant aspect of the total capital cost in most cases studied and so this change could result in considerable differences in the economic assessments undertaken. As such it is difficult to draw definitive conclusions and the recommendation is that the cases are modelled to confirm performance as per literature with bottom up estimates made for each case using a single source of data. This will ensure that the results are developed using the same basis and methodology and are therefore directly comparable. However this was beyond the scope of this present study. However, this result and the result for Case 1 of a cost of CO_2 avoided of 33.5 €/t CO_2 does, however, illustrate the potential for retrofit opportunities utilising a FC hybrid scheme.

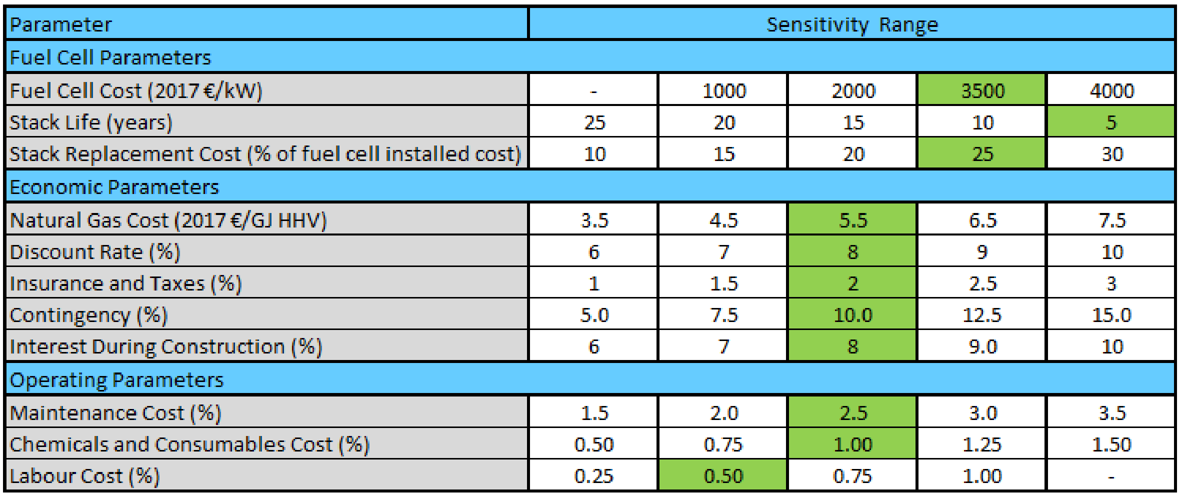
As Case 1 gives the best overall cost performance for the CO_2 capture cases it has been used as the base case in the sensitivity analysis. However, the general findings of the sensitivity study could be applied to all of the FC cases considered.

5.19. Sensitivity Analysis

The results presented in Table 12 are based on a number assumptions and values taken from the literature review. Some parameters, such as the cost of the FC, had a wide variation in the studies considered and a sensitivity analysis has been carried out in three key areas to determine the relative effects.

The areas considered are FC parameters, economic parameters and operational parameters. The parameters and range of values used in the sensitivity analysis are presented in Table 13.

Table 13: Case 1 Atmospheric NGFC-MCFC with CO_2 Capture: Sensitivity Study Parameters



Parameter	Sensitivity Range				
Fuel Cell Parameters					
Fuel Cell Cost (2017 €/kW)	-	1000	2000	3500	4000
Stack Life (years)	25	20	15	10	5
Stack Replacement Cost (% of fuel cell installed cost)	10	15	20	25	30
Economic Parameters					
Natural Gas Cost (2017 €/GJ HHV)	3.5	4.5	5.5	6.5	7.5
Discount Rate (%)	6	7	8	9	10
Insurance and Taxes (%)	1	1.5	2	2.5	3
Contingency (%)	5.0	7.5	10.0	12.5	15.0
Interest During Construction (%)	6	7	8	9.0	10
Operating Parameters					
Maintenance Cost (%)	1.5	2.0	2.5	3.0	3.5
Chemicals and Consumables Cost (%)	0.50	0.75	1.00	1.25	1.50
Labour Cost (%)	0.25	0.50	0.75	1.00	-

For each sensitivity case the parameter was adjusted through the sensitivity range while holding all other parameters as per the base case, Case 1 Atmospheric NGFC-MCFC with CO_2 Capture (highlighted in green in Table 13). The LCOE was then determined using the same methodology as described above and plotted on a tornado chart to show the effect of each parameter on the LCOE, as presented in Figures 55, 56 and 57 and **Error! Reference source not found.**.

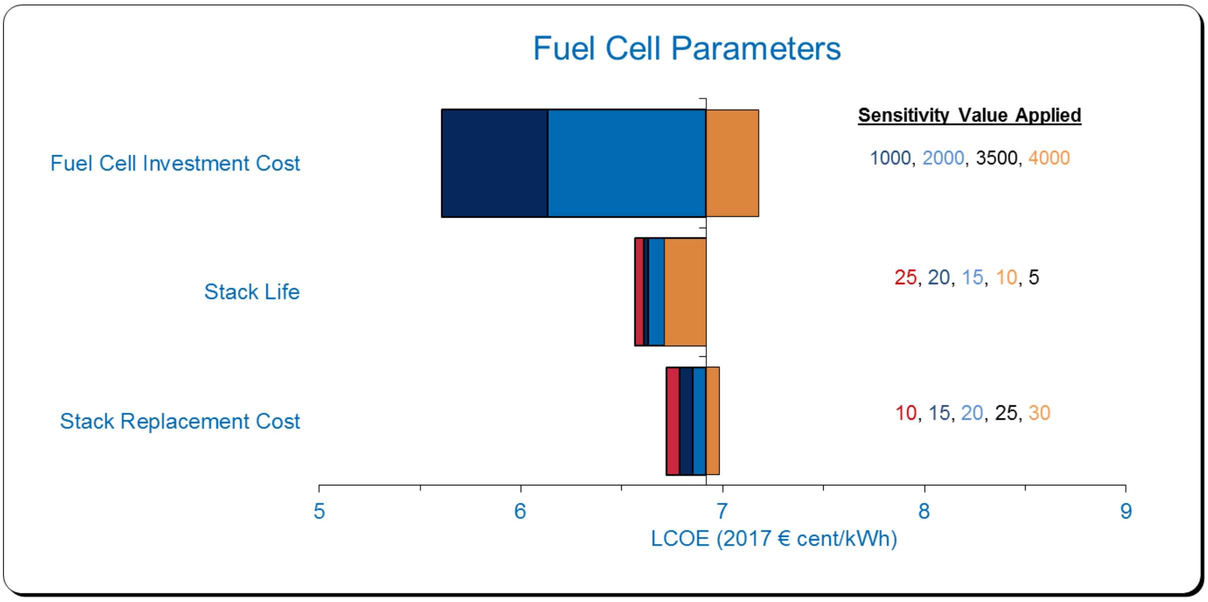


Figure 59: Case 1 Atmospheric NGFC-MCFC with CO_2 Capture: LCOE Sensitivity to FC Parameters

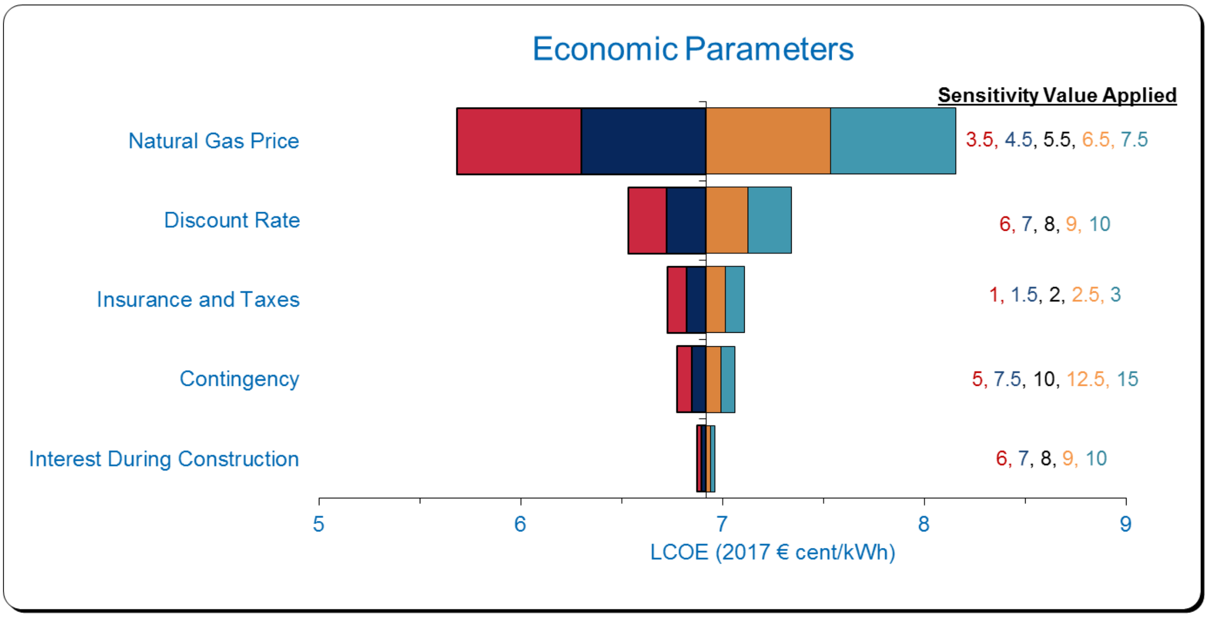


Figure 60: Case 1 Atmospheric NGFC-MCFC with CO_2 Capture: LCOE Sensitivity to Economic Parameters

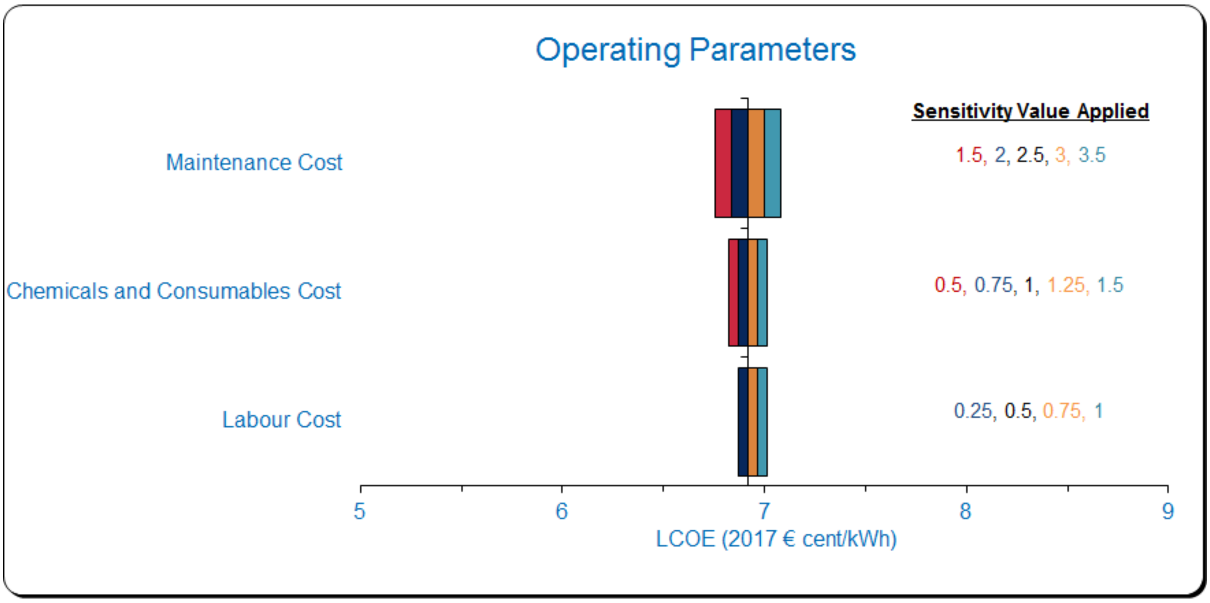


Figure 61: Case 1 Atmospheric NGFC-MCFC with CO_2 Capture: LCOE Sensitivity to Operating Parameters

Table 14: Sensitivity Study Results Summary

Fuel Cell Cost (2017 €/kW_e)		1000	2000	3500	4000
Calculated LCOE (2017 € cent/kWh)		5.61	6.13	6.92	7.18
Stack Life (years)	25	20	15	10	5
Calculated LCOE (2017 € cent/kWh)	6.56	6.61	6.63	6.71	6.92
Stack Replacement Cost (% of fuel cell installed cost)	10.00	15.00	20.00	25.00	30.00
Calculated LCOE (2017 € cent/kWh)	6.72	6.79	6.85	6.92	6.98
Natural Gas Cost (2017 €/GJ HHV)	3.5	4.5	5.5	6.5	7.5
Calculated LCOE (2017 € cent/kWh)	5.68	6.30	6.92	7.54	8.15
Discount Rate (%)	6.00	7.00	8.00	9.00	10.00
Calculated LCOE (2017 € cent/kWh)	6.53	6.72	6.92	7.13	7.34
Insurance and Taxes (%)	1	1.5	2	2.5	3
Calculated LCOE (2017 € cent/kWh)	6.73	6.82	6.92	7.01	7.11
Contingency (%)	5	7.5	10	12.5	15
Calculated LCOE (2017 € cent/kWh)	6.77	6.85	6.92	6.99	7.06
Interest During Construction (%)	6.00	7.00	8.00	9.00	10.00
Calculated LCOE (2017 € cent/kWh)	6.88	6.90	6.92	6.94	6.96
Maintenance Cost (%)	1.5	2	2.5	3	3.5
Calculated LCOE (2017 € cent/kWh)	6.76	6.84	6.92	7.00	7.08
Chemicals and Consumables Cost (%)	0.50	0.75	1.00	1.25	1.50
Calculated LCOE (2017 € cent/kWh)	6.82	6.87	6.92	6.97	7.01
Labour Cost (%)		0.25	0.50	0.75	1.00
Calculated LCOE (2017 € cent/kWh)		6.87	6.92	6.97	7.01

The largest effects on the LCOE performance can be seen in sensitivities for FC investment cost and natural gas fuel price. Reducing the FC investment cost from 3500 €/kW_e to 1000 €/kW_e results in a LCOE of 5.61 € cent/kWh, an 18.9% reduction when compared to the base case, Case 1.

An increase in the natural gas price from 5.5 to 7.5 €/GJ (HHV) would result in an LCOE of 8.15 € cent/kWh, a 17.8% increase when compared to the base case.

In the literature review, degradation was identified as a key area to improve the commercial viability of FC hybrid cycles. To model sensitivities in degradation the base case degradation rate of 0.6% per 1000 hours was adjusted with the stack life increased accordingly due to the fact that, with less degradation, the frequency of stack change would be decreased. For example, the sensitivity case that has a stack life of 10 years (twice the stack life of the base case) is modelled with a degradation rate of 0.3% per 1000 hours and results in an LCOE of 6.71 € cent/kWh, a 3.0% decrease when compared to the base case.

The other parameters considered have less of an impact (up to 5.6% change in LCOE) and, as many of them are percentages linked to the capital cost, the changes in LCOE are proportional to the sensitivity step changes made to each parameter.

As a further sensitivity, the effect of assuming a fixed operating cost rather than a percentage linked cost was investigated. The fixed operating costs are defined as the Maintenance, Labour and Chemicals and Consumables. For maximum sensitivity effect the lowest fixed operating costs, €35.2M/year for Case 1, were applied to Case 4 (Pressurised NGFC-SOFC with CO_2 Capture), the case with the highest installed cost and hence highest fixed operating costs of €148.2M/year. The revised fixed operating costs results in a LCOE for Case 4 of 17.22 € cent/kWh, a 11.9% reduction in LCOE.

6. DISCUSSION

Case 1 (Atmospheric NGFC-MCFC with CO_2 Capture) and Case 5 gave the best performance in terms of LCOE and cost of CO_2 avoided compared to all the CO_2 capture cases. Appendix 3 Table A3-1 illustrates a number of similarities between these cases and differences to the other cases studied.

Whilst Case 1 and Case 5 have lower CO_2 capture rates of 72% and 88% respectively compared to >90% for the other CO_2 capture cases considered, it is anticipated that even with enhanced CO_2 capture both cases would still reflect the best LCOE performance particularly when both cases demonstrate the best cost of CO_2 avoided.

Both Case 1 and Case 5 employ MCFCs in hybrid cycles, utilising combustion flue gases from NG and PC plant respectively as the cathode feed and natural gas at the anode feed. Employing the MCFC in this configuration results in the MCFC process concentrating CO_2 at the anode exhaust, from which heat can be recovered in a steam cycle and CO_2 removed through cryogenic CO_2 separation or oxy-combustion followed by combustion. This plant configuration can capture 75% of CO_2 with a negligible efficiency variation, while increasing the overall power output by about 20% (Campanari, et al., 2014). Hence there is a net plant performance benefit in capturing CO_2 albeit with the significant economic penalty associated with the FC cost. Higher levels of CO_2 capture and larger increases in additional power can be achieved by increasing the amount of fuel cells used.

Of particular interest is the potential for retrofit of MCFC systems to NG and PC plants to enable carbon capture at a lower efficiency penalty than current state-of the-art amine scrubbing carbon capture with a lower cost for CO_2 avoided. Spinelli, et al. (Spinelli, et al., 2018) found that their proposed MCFC retrofit configurations applied to PC and NG plants yields a limited (<3 percentage points) decrease in efficiency, compared to ~6 to 8 percentage points efficiency penalty for amine scrubbing (NETL, 2015). Spinelli, et al. also found a cost for CO_2 avoided in the range 25 to 40 €/tonne. This shows good agreement with the results of our analysis with a cost of CO_2 avoided of 33.5 €/tonne for Case 1 which is competitive with state-of-the-art amine scrubbing carbon capture.

Case 2 (Atmospheric IGFC-SOFC with CO_2 Capture), Case 3 (Pressurised IGFC-SOFC with CO_2 Capture) and Case 4 (Pressurised NGFC-SOFC with CO_2 Capture) consider a SOFC in hybrid cycles utilising either syngas produced from coal gasification or natural gas as a fuel. In these cases, the larger proportion of electricity is produced by the FC which has higher power production efficiency than the equivalent gas or steam turbine. Compared to conventional power flue gases, the CO_2 in the exhaust gas from a FC is easier to capture through simpler processes due to the fact it predominantly consists of CO_2 and H_2O. Both of these factors mean that the overall cycle efficiency of a FC hybrid cycle with CCS, is greater than that of a conventional NGCC or PC plant with CCS, however, , there is a significant economic penalty in terms of the cost of the FC at present.

In both Case 1 and Case 5, the FC power contribution to the overall net power output for the hybrid cycle scheme is relatively small, 19% and 33% respectively, compared to the 94%, 74% and 109% for Cases 2, 3 and 4. This means that in Case 1 and Case 5 a greater proportion of the plant power burden is met by mature technologies (i.e. gas and steam turbines) that have a lower cost per kW_e of power output. Therefore the FC price of 3500 €/kW_e has less of an impact on the LCOE of these cases than in Cases 2 to 4. Given this result it would prove beneficial to investigate the optimum balance between FC power and CO_2 capture rate in terms of LCOE performance. As the price of a FC decreases towards the lowest FC cost of 1000 €/kW considered in the sensitivity study the LCOE for Cases 2 to 4 approaches that of NGCC plus CCS.

Fuel Cell Investment Cost

Throughout the literature review, and as highlighted in the sensitivity study presented here, FC cost is cited as the major barrier to commercialisation of FC plus CCS hybrid cycles. The key to a reduction in the cost for the FC is often linked to an increase in demand and hence volume of production, with the resultant savings driven through manufacturing efficiencies such as higher utilisation of equipment and labour. A major challenge with this is that for an increase in production to occur then the FC has to be at a price that makes the process economically attractive. Based on the results of the study a reduction in FC cost from 3500 €/kW$_e$ to around 2000 €/kW$_e$ would be required to drive demand from an economics perspective assuming policy, fuel price etc. remain as is. Attempts to reach this FC cost should be made through further improvements in material development and manufacturing techniques where possible.

The fuel cell cost of 3500 €/kW$_e$ was selected based on the average FC costs observed in the literature. It is noted that there are some limitations in this approach in that many of the studies consider small-scale, FC systems which have higher costs than are anticipated for large scale power production. However, given that there is a lack of large scale demonstrations of this type of fuel cell application this somewhat conservative approach was considered acceptable with an appropriate sensitivity analysis carried out with cost reductions applied.

Fuel Cell Stack Life

Another key finding from the literature review was a requirement to improve the current degradation rates of both SOFCs and MCFCs and hence FC stack life. For the purpose of the economic assessment a degradation rate of 0.6% per 1000 hours was used resulting in a “useful” cell stack life of 5 years. After 5 years operation it is considered more economical to replace the stack given the level of degradation and resultant impact on FC performance. The sensitivity study considered a stack life up to 25 years by means of an adjustment in the degradation rate. As can be seen in Figure 59 stack life does have an effect on the LCOE but greater reductions can be seen in addressing FC cost. Indeed a lower FC cost should result in a lower stack replacement cost which in turn should allow for a greater degradation rate while remaining economic i.e. the reduced FC cost allows for more frequent stack changes. That is not to say that improvements in FC degradation rate should not be targeted, but a greater emphasis should be placed on reducing FC costs to have the most significant impact on the LCOE, and hence economic viability of FC plus CCS hybrid cycles.

Pressurised SOFC / MCFC Configuration

Case 2 and Case 3 highlight the effect of pressurisation albeit with a number of differences in the plant technology (post-anode oxy-combustor vs oxidation reactor respectively).

In comparison to atmospheric Case 2, pressurised Case 3 delivered a lower LCOE and this mirrors the findings of NETL (NETL, 2009) which showed an improved LCOE with increased operating pressure. An increase in pressure results in an increase in FC efficiency for both SOFCs (Henke, et al., 2012) and MCFCs (Duan, et al., 2016) as shown in Figure 62 and Figure 63 respectively. These efficiency improvements contribute to the improved LCOE performance of pressurised FC systems, although it is tempered somewhat by increased system complexity and additional installed cost requirements through increases in auxiliary equipment such as gas compressors, pressure vessels and the use of more exotic materials. For FC plus CCS hybrid cycles, the FC operating pressure is a key variable that should be used for optimisation by consideration of the whole system.

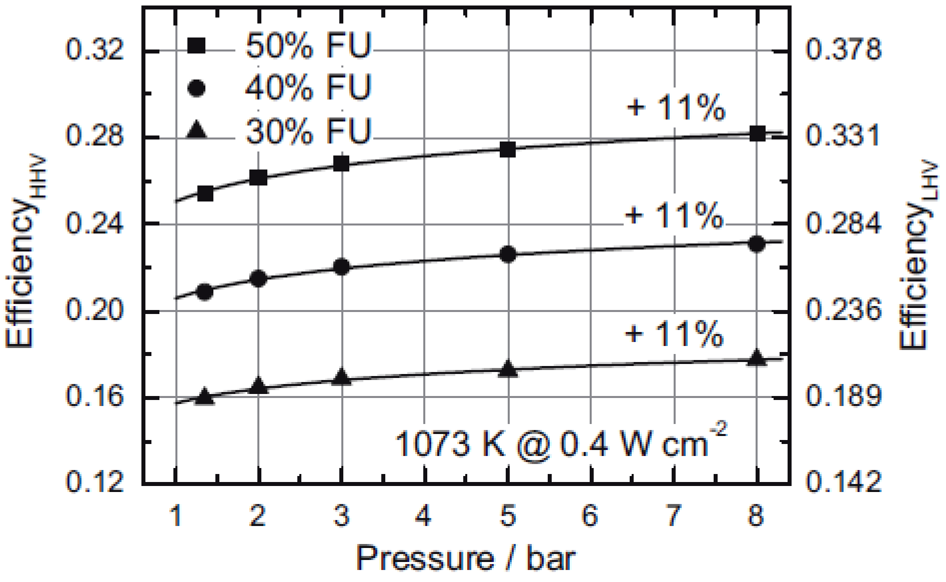


Figure 62: Effect of Pressure on SOFC Efficiency for Different Fuel Utilisations at Constant Power Density (Henke, et al., 2012)

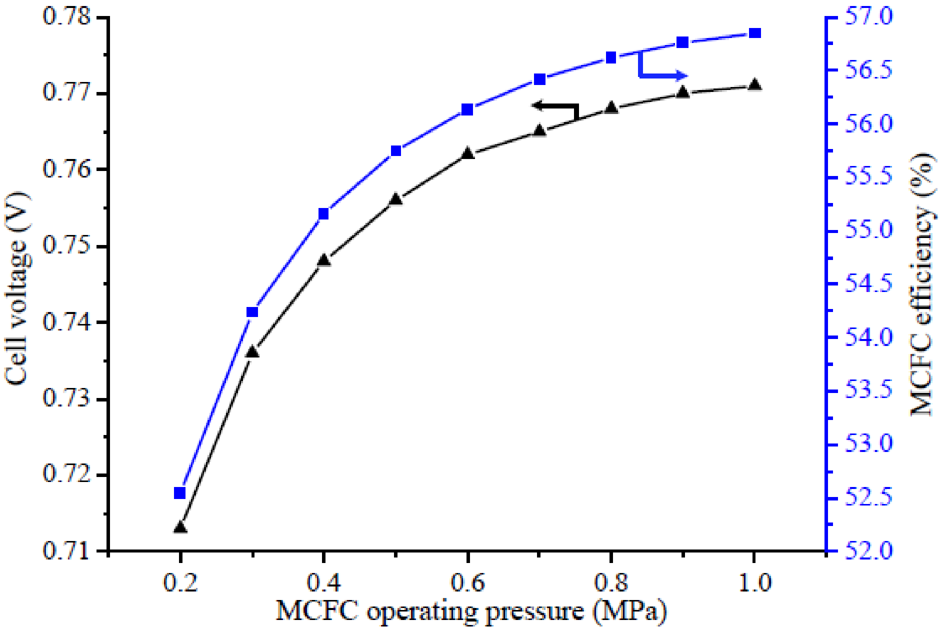


Figure 63: Effect of Pressure on MCFC Cell Voltage and Efficiency (Duan, et al., 2016)

Concluding Remarks

The results of this study show that FC with CCS hybrid cycles have the potential to be competitive with, or even surpass, current state-of-the-art carbon capture technology but, as of yet there are no large scale demonstrations which would be required to increase confidence in the technology and reduce process risk. In addition, other emerging technologies, such as the Allam Cycle that reports an LCOE of approximately 60 US$/MWh and a HHV cycle net efficiency of 48% for coal and 53% for gas (Lu, 2017), could act as competitors to FC CO_2 capture systems.

As it stands, state-of-the-art amine scrubbing technology is still the likely choice for carbon capture due to the state of development and demonstration at scale (Boundary Dam, Petra

Nova). However, the 50 MW_{th} Net Power Allam Cycle demonstration plant in La Porte, Texas and FuelCell Energy's modified 2.8 MW_e SureSource 3000 MCFC CO_2 capture demonstration at James M. Barry Electric Generating Plant in Alabama, capturing CO_2 from NG and PC flue gas respectively, are due to be operational in 2018. These demonstration projects will provide key performance data to enable further development of the next generation of carbon capture plants and increase confidence in the respective technology options.

7. CHALLENGES AND BARRIERS

7.1. Manufacturing & Materials

At the time of writing, FCs systems are on average still 3-4 times more expensive in terms of capital investment than conventional distributed generation, as seen in Table 15.

Table 15: Comparison of FC systems with other engines

	Diesel Engine	Gas Engine	Gas Turbine	Steam Turbine	FCs
Investment cost (US$/kW)	1,100-1,300	1,100-1,300	2,000-2,500	1,100-1,300	3000-5000

Fuel cell costs can be broken into three areas: the material and component costs, labour (i.e. design, fabrication, and transport), and capital cost of the manufacturing equipment (Wang, 2015). It should be mentioned that only labour and capital costs can be reduced through mass-manufacturing. Material and component costs, such as catalysts, membrane and bipolar plates, are dependent on technological innovations and the market, however, reduction of system manufacturing costs is certainly expected with increasing mass production.

In the large-size stationary market, economies of scale in manufacturing, simplifications in fabrication methods, and technological developments should be able to reduce selling prices in future, but the extent of price reductions may be limited. The increasing average size of installations should fuel significant growth for the coming decade, but given price uncertainties growth beyond that becomes less certain. A market projection study for MCFC systems has been undertaken by ORNL, suggesting that by 2025, MCFC systems could be entering the US market at approximately 340 MW_e/year. SOFC systems are also likely to become important in this size range, having experienced significant percentage-wise growth in recent years. Accordingly, the manufacturing costs of fuel cell stacks are improving rapidly, driven primarily by improvements to power density. Regarding SOFC, the ‘mature' manufacturing costs have decreased by an order of magnitude over the past decade, from 1500 US$/$kW_e$ in 2000 to between 44 and 175 US$/$kW_e$ in 2013, resulting in a power density between 0.5 W/cm^2 and 2 W/cm^2 respectively (Adams, et al., 2013).

Potential cost increases in the prices of steel, nickel, and zirconia could seriously affect the economics of the stack. However, stack construction improvements, such as thinner interconnects and anodes, as well as manufacturing process improvements can help to offset a potential increase in raw materials costs. Further cost reductions to the balance of plant (heat exchangers, compressors, etc.), as well as interconnects, seals, and manifolds can be achieved when lowering SOFC / MCFC operating temperatures since lower-cost materials and manufacturing methods may be used at those conditions. Interconnects contribute to a particularly large portion of stack costs and materials, which creates a scale-up challenge since larger stacks will require more interconnects.

The materials which comprise the cell stack are another major contributing factor to the performance of SOFCs and MCFCs. Anode materials must satisfy certain requirements of catalytic activity, electronic conductivity, thermal capability, chemical stability, porosity, carbon deposition, and sulphur poising. Similar criteria exist for cathodes, electrolytes, seals, and interconnects. Finding affordable materials (or ways of using less) to meet these performance requirements is an area of active research. Yttrium, lanthanum, and cerium are typically the most crucial contributors to manufacturing of the cells, however, price, access, or global reserves of these particular materials are currently not an issue that prevents commercialisation of those particular fuel cells. For example, although yttrium in particular has skyrocketed in price (moving from 5 to 160 US$/kg from Yr2006 to Yr2011), it remains less than 1% of the total installed cost of SOFC stacks (Thijssen, 2011), since most of the material in stacks consists of the interconnects between them. Lastly, mass commercialisation of SOFCs / MCFCs is not expected to put significant pressure on rare earth element markets by itself. As a result, commercialisation challenges related to materials will likely be limited to the development and fabrication of improved materials, rather than supply chain issues.

7.2. Operational challenges

7.2.1. Pressure Management

One of the challenges of pressurised fuel cell configurations is to manage the safe operation at high pressures.

In this instance fuel cell stacks need to be placed inside a pressure vessel, adding to the cost and complexity of the system. Furthermore, there is a risk of cell destruction / damage when there is a significant pressure difference between the anode and cathode side, causing the cell to break or crack. Effective control is required to prevent significant pressure changes arising from disturbances or transient operation which can cause damage. However, several fuel cell stacks have demonstrated successful operation at high pressure, for instance Rolls Royce (LG) (6.4 bar), the German Aerospace Centre (8 bar), the Korea Institute for Energy Research (3.5 bar), Westinghouse (15 bar) and the Chinese Academy of Sciences (6 bar) (Adams, et al., 2013).

Complex but precise pressure control has been demonstrated in maintaining pressure differences as low as 10 mbar. Metal-supported cell stacks may be particularly beneficial for pressurised applications due to their ruggedness and mechanical strength and may even be able to support large pressure differences between anode and cathode. Other systems integration challenges due to pressurised operation include issues relating to backflow, compressor surge, and compressor stall, especially for the SOFC-GT hybrid systems.

7.2.2. Oxygen Handling

As discussed previously under Section 2.2, if post-anode CCS using condensation-based techniques is preferred, the unreacted fuels leaving the anode need to be oxidised. This is required because the pipelines which carry CO_2 are currently subject to purity limitations not only on CO_2 purities but also other contaminants. Therefore, air cannot be used as the oxygen source, instead, high-purity oxygen is required.

Cryogenic distillation - air separation - techniques are commonly used to produce high purity O_2 for gasification and other purposes and can be used to provide O_2 for post-anode oxidation as well. However, stoichiometric amounts of O_2 are required. High O_2 levels would stress the very tight O_2 bounds of the CO_2 pipeline. Conversely low O_2 levels would result in unreacted fuel which could violate the CO_2 or hydrocarbon limits and also lead to reduced waste heat produced for use in the integrated heat recovery systems. This creates a systems integration challenge of detecting and responding to disturbances in flow rate, conversion in the FC and the O_2 produced by air separation.

A reliable control system will need to be developed in order to handle this type of system integration before second generation SOFC systems integrated with CO_2 capture can be commercialised. O_2 production using other technologies such as ceramic auto thermal recovery (Shelley, 2009) and ionic transport membranes could potentially result in energy and cost advantages (Kneer, et al., 2010), however, these technologies are not yet fully available for large scale applications.

Another solution would be to integrate the oxidising step with the air supply preheater in one unit. This way the heat generated by oxidation is removed by preheating cool air for the SOFC cathode. This integrated approach could prevent excessive temperatures in the oxidiser from being developed and simplifies issues with materials and systems integration associated with high temperatures. An integrated catalytic oxidiser and heat exchanger has been successfully demonstrated at laboratory scale and the approach is expected to contribute to significant cost reductions if applied to large-scale systems in the future (Jolly, et al., 2012).

7.2.3. Limitations of FC Performance

The fuel cell performance is affected by various parameters including current density, gas composition, fuel utilisation, pressure and temperature, as discussed previously. When the operating conditions for a fuel cell are changed the system cost may increase. Ideally, the fuel cell system must combine high power density at low cost. For stationary power applications, higher voltages and hence higher efficiency are required.

Figure 64 presents the voltage–power relationship for a fuel cell. It is clear that the higher current and power density usually occur for low cell voltage. Hence, for maximum efficiency, the operating conditions should be balanced to get a good compromise between current density, voltage and operating cost.

For example, increasing the SOFC working voltage results in more efficient operation, however, it also results in a decrease in the average current density. This means that for a constant total power output more SOFC stacks would be needed, resulting in higher capital cost. It should also be noted that due to the theoretical limit of the Nernst potential, there is a certain constraint of further increasing the SOFC working voltage

Another major factor that determines fuel cell efficiencies is gas composition and utilisation of reactants. The fuel utilisation is given by:

$$U_{fuel} = \frac{Fuel_{in} - Fuel_{out}}{Fuel_{in}}$$

It is apparent, if the fuel and oxygen gas have a higher concentration, then the fuel cell will possess higher efficiencies and hence high cell voltages. The difference between fuel provided at the inlet and fuel flow rate at the outlet of the FC represents the fuel consumed. Other factors such as fuel leakage and chemical reaction may lead to higher fuel consumption, leading to an increase of fuel utilisation. From inlet to outlet, the gas composition keeps changing, hence leading to reduced cell voltages. The voltage reduction is attributed to the lowest electrode potential (from the Nernst potential described in Section 3.1) because the electrodes are good ionic conductors and the cell voltage values cannot be greater than the minimum Nernst potential.

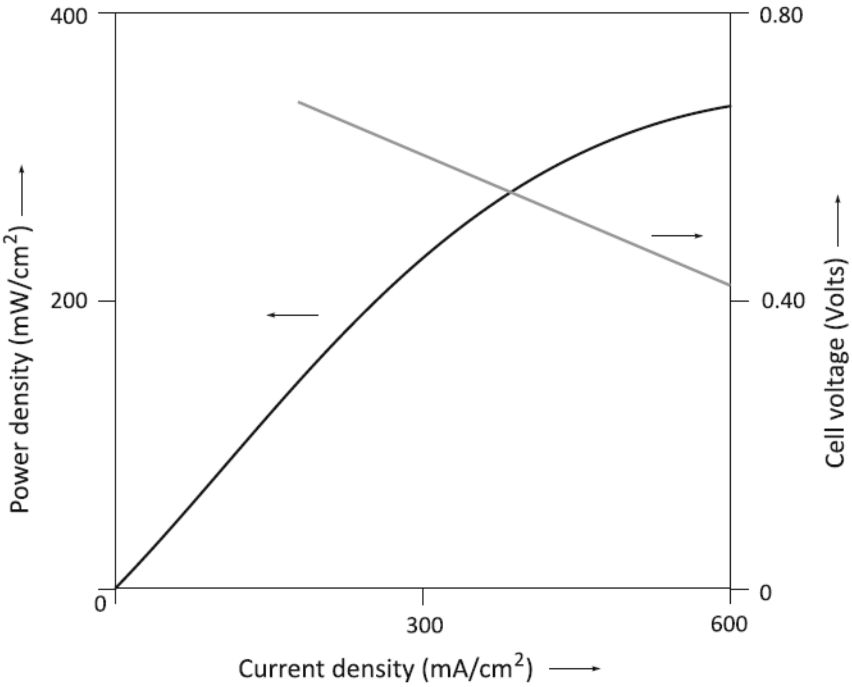


Figure 64: The relation between current density, power density and cell voltage of fuel cell

7.3. Public Policy

Whilst fuel cells have been studied extensively for the past 70 years, most technological advancements have occurred in the last two decades due to an increased global interest in the use of hydrogen as a fuel. This interest was generated by an increased desire to improve energy efficiency in power generation for environmental, cost, and energy security reasons. However, manufacturing challenges contributing to high capital costs of fuel cells must be overcome before widespread commercialisation can take place. SOFC and MCFC costs are decreasing rapidly as these challenges are being addressed at the R&D level. Increase of production is also expected to result in capital cost reduction due to the economies of scale.

However, the fuel cells market – and especially the CO_2 capture integrated fuel cells - can be significantly influenced by environmental public policies or even hints of future policies related to CO_2 emissions and fuel conservation. Public policy can influence the market by placing restrictions on emissions in the form of limits, taxes, or credits; specifically favouring via taxation, public funding or subsidies or discouraging the development of certain technologies. As these policies begin to take shape, the concept of fuel cells combined with CO_2 capture becomes increasingly attractive since the FCs do not only have higher electrical efficiencies in the power cycle, but also could potentially facilitate CO_2 capture at lower costs than conventional forms of fossil-fuel-based power generation, as seen in this study. Thus, the influences of public policies can add market value to FCs in the form of an avoided cost of CO_2 emissions.

Given their early stage of commercialisation, fuel cell systems usually still require financial support. Support mechanisms come in the form of grants (e.g. Japan), enhanced depreciation allowances for commercial applications (e.g. Japan, USA, UK, Germany, Italy), feed-in tariffs (e.g. UK), and fuel price discounts (e.g. Germany). The level of support provided by these schemes varies greatly, with marked impact on uptake in each country. Examples of successful programmes include the grant systems in Japan, and the tax credit system in the USA. In the short-to-medium term, fuel cell systems will require some form of public support to stay on the path to standalone competitiveness. At the time of writing the fuel cell industry appears to be going through a maturing phase with focus on manufacturing methods, quality control, safety systems, system control and balance of plant (as opposed to fundamental R&D for stack design). Continued support through this period will be critical.

Coal-based systems using SOFCs for bulk power generation are predicted to be significantly more economical than PC and IGCC when CCS is required and even more economical than with traditional NGCC (without CCS) for natural gas prices higher than 6.5 to 6.75 US$/MMBtu (Adams, et al., 2013). Consequently, if these regulations take effect, integrated SOFC systems with CCS will become one of the few economically feasible power generation options for coal. Secondly, SOFCs could also address the problems created by uncertainties about future regulations. Since the CO_2 capture, compression, and storage processes are entirely "downstream" from the power generating portions of the plant, a SOFC-based power generation system without CCS technology can be constructed in the near term and at a later stage, if regulations for CCS remain favourable, add a CCS retrofit section with minimal impact on the main process, low parasitic power demands, and comparatively low cost. PC power generation does not have this advantage since the integration of a CCS unit could cause LCOE to almost double and reduce the net power output of the plant significantly (Woods, et al., 2012). Retrofitting CCS to IGCC is less expensive and it may become even more favourable in the future in EU countries that have started switching to the H_2 economy, shutting down the large centralised PC power plants.

SOFCs and MCFCs are also commercially advantageous for power generation companies that are not tied to a particular type of fuel because both fuel cells are flexible with regard to feedstock. A company using SOFCs or MCFCs is prone to less risks of loss if one particular type of fuel becomes too expensive or even prohibited, since the fuel can be switched (with the appropriate processing and treatment steps). For example, the current low price of natural gas is mainly driven by large shale gas discoveries coupled with breakthrough advances in low cost hydrofracking. Saving in fuel costs due to the improved efficiencies that the SOFCs / MCFCs can offer is less meaningful when the price of fuel is low. In that case, the lifetime cost differences between SOFC and traditional technologies favours fuel cells less, compared to traditional combustion technologies. However, the future of fuel price and supply remains uncertain; while natural gas is being touted as a near-term way to reduce CO_2 emissions quickly (by displacing coal), it is quite possible that natural gas will eventually be targeted for replacement by renewables such as biomass (which can be gasified for use in SOFCs and MCFCs). Thus the fuel-flexibility of SOFCs and MCFCs can be quite advantageous.

Although the potential integrated fuel cells with CO_2 capture systems discussed in this study appear promising, there are many challenges that must be overcome before widespread commercialisation can take place. These include the difficulties in manufacturing, scale-up, materials, cell operational issues, system integration issues, propensity for cell failure due to rapid temperature transients, degradation from anode oxidation, and unreliable seals (Adams, et al., 2013). All of these contribute to higher investment costs and/or operating costs by affecting the cost of manufacturing, the cost of support auxiliaries such as control systems, and/or the lifetime of the unit, resulting in high LCOE.

7.4. Deployment of Large-scale CCS

Deployment of CCS projects worldwide is facing many challenges, including technical and financial issues, public acceptance and the establishment of regulatory frameworks (International Energy Agency (IEA), 2013).

For the most part, CCS has not progressed beyond the demonstration stage, with stakeholders hesitant to commit the resources necessary to scale-up and roll-out deployment of the technology. This hesitant progress reflects several characteristics: the large financial magnitude of individual investments, the creation of new networks, the relatively high level of perceived technology and real commercial and policy risks, together with the often tentative commitments of governments to deep decarbonisation investments.

One factor that could directly influence governmental policy mechanisms is carbon pricing, which appears insufficient to incentivise and deliver CCS commercially at the moment (Element Energy & Vivid Economics, 2018). A high future carbon price seems too distant and uncertain to incentivise the near term CCS development that is necessary to prepare the way for widespread deployment in the future. In addition, a carbon price only addresses the externality of CO_2 emissions, whereas there are a number of other market failures, such as counterparty risk and natural monopolies, which also merit intervention.

Various regulatory approaches are under development in most countries that have significant potential CO_2 storage resources and CCS activities. Despite the approach taken, it should be ensured that their regulatory framework is kept up to date with the accumulation of new knowledge regarding CCS. The lack of a robust and comprehensive regulatory framework creates an environment of uncertainty that slows down the progress of CCS demonstration projects.

Lastly, understanding of the value proposition of large scale deployment of CCS is key. In particular, governments may wish to determine the future role of CCS within their own decarbonisation trajectories and explore how investments now might avoid larger decarbonisation costs later (associated with deployment of less cost-effective technologies). The role of CCS in avoiding large decarbonisation costs in the long term is a key aspect of the value of CCS, particularly from the point of view of consumers. CCS enables access to large amounts of low carbon energy, reducing prices of a large amount of consumer goods compared to a decarbonisation scenario with no CCS.

8. CONCLUSIONS

Using the information gathered in the literature review and the techno-economic analysis it was concluded that:

8.1. Fuel Cell Technology

- Fuel Cell technologies with CO_2 capture is a promising alternative to conventional technologies and has a role to play in the next generation CCS.
- A number of studies can be found in the literature, focusing on SOFCs and MCFCs integrated in power cycles. However, it is evident that SOFC configurations have been the main focus recently. This may be due to the fact that SOFC technology has significant room for technical, hence economic, improvement. On the other hand, for MCFCs, the FC stack accounts for approximately 60% of the MCFC system costs (BOP, gas clean up, pre-reformer, water management, heat exchangers, control, inverter) (NREL, 2010). The same study concludes that no single issue could achieve significant cost reductions, but stack life time, power density and cost reduction of gas cleaning are the most important R&D areas to bring down overall system costs. However, even under the most optimistic circumstances, it is not likely that first costs for an MCFC power plant can be brought much below 2,000 US$/kW$_e$ (NREL, 2010).
- Several methods of CO_2 capture / separation can be employed together with FCs for further CO_2 purification:
    - Cryogenic CO_2 separation
    - Oxy-combustion
    - Oxidation
    - Condensation
    - Membrane Separation
    - Physical Absorption
- Of all the CO_2 capture methods, cryogenic CO_2 separation and oxy-combustion with condensation are the most widely considered.
- There are several factors that affect the performance of SOFC and MCFC, including:
    - Pressure
    - Temperature
    - Voltage
    - U_f
    - U_{Air} for SOFC and U_{CO2} for MCFC
- Optimum operational and plant parameters of the FC depend on individual design configurations. Appendix 2 details some of the key plant areas and options for consideration.
- Pressurised configurations showed significant potential with regard to the overall power plant performance. At a constant current density, increases in pressure result in an increase in voltage. However, with increasing pressure the voltage increases get smaller (Duan, et al., 2011). Therefore there is a diminishing return in pressure increases and eventually the increases in performance are outweighed by the significant costs and complexity of the system caused by pressurisation. For a given system the optimum operating pressure should be determined.

8.2. Techno-Economic Performance

- The techno-economic analysis showed that fuel cell systems plus CO_2 capture cases have the potential to be competitive with the current Cansolv based scrubbing system.

However, limitations in the techno-economic analysis are noted. The cases taken from literature and updated in this present study are all based on simulated performance data and not on performance data of real plants. Therefore, validation of FC performance against the reference curves utilised would be required to determine the effects on the simulated overall system performance.

- Techno-economic Case 1, a hybrid cycle consisting of a NGCC power plant and an atmospheric pressure MCFC equipped with cryogenic carbon capture technology, gave the best economic performance of all CO_2 capture cases, with an LCOE of 6.92 € cent/kWh. This represents a 23.9% decrease on the equivalent NGCC plus Cansolv PCC Reference Case 2b.
- The cost of CO_2 avoided is 33.5 €/t CO_2 relative to NGCC Reference Case 2a for techno-economic Case 1. This represents a 65.1% decrease on the equivalent NGCC plus Cansolv PCC Reference Case 2b.
- The sensitivity analysis showed that the fuel cell investment cost was the key parameter affecting the LCOE for Case 1.
- Throughout the literature review, and as highlighted in the sensitivity study, FC investment cost is cited as the major barrier to commercialisation of FC plus CCS hybrid cycles. The key to a reduction in the cost for the FC is often linked to an increase in demand and hence volume of production, with the resultant savings driven through manufacturing efficiencies such as higher utilisation of equipment and labour. A major challenge is that for an increase in production to occur then the FC has to be at a price that makes the process economically attractive. In addition to cost reductions driven by R&D, time and scope-limited subsidies could be used as a driver for market introduction (Fuel Cells and Hydrogen Joint Undertaking, 2015).
- Based on the results of the study a reduction in FC investment cost from 3500 €/kW$_e$ to around 2000 €/kW$_e$ would be required to drive demand from an economics perspective assuming policy, fuel price etc. remain as is. In addition to savings anticipated through increases in production volumes, attempts to reach this FC cost could potentially be made through further improvements in material development where possible, improved leaner and automated manufacturing techniques, simplified designs and an engaged supply chain (Fuel Cells and Hydrogen Joint Undertaking, 2015).
- Enhancement of the current degradation rates in both SOFCs and MCFCs were identified as another key area to achieve a lower cost. However, the sensitivity study showed that, although important, the FC degradation rate effects were not as substantial as the effect of FC cost. Improvements in degradation rate should be targeted, but a greater emphasis should be placed on reducing FC costs to have the most significant impact on the LCOE, and hence economic viability of FC plus CCS hybrid cycles. The US Department of Energy’s Solid State Energy Conversion Alliance (SECA) coordinates Federal efforts to facilitate development of a commercially relevant and robust SOFC system. The key SECA objectives are an efficiency of greater than 60 percent, meeting a stack target cost of 175 US$/kW$_e$ and demonstrating lifetime performance degradation of less than 0.2 percent per 1,000 hours over a 40,000 hour lifetime. This emphasises the importance of both the cost and degradation improvements to achieve a commercial offering.
- Techno-economic Case 1 Atmospheric NGFC-MC with CO_2 Capture and Case 5 Atmospheric Hybrid-MCFC with CO_2 Capture both employ MCFCs utilising the flue gas from NG and PC processes respectively as a feedstock for the cathode. This is of particular interest, as it gives the potential for MCFC systems to retrofit NG and PC

plants at lower efficiency penalty than current state-of-the-art amine scrubbing carbon capture, potentially with a lower CO_2 avoidance cost.

- Spinelli, et al.(2018) found that proposed MCFC retrofit configurations applied to PC and NG plants yielded a limited (<3 percentage points) decrease in efficiency, compared to ~6 to 8 percentage points efficiency penalty for amine scrubbing (NETL, 2015). Spinelli, et al. also found a CO_2 avoidance cost in the range 25 to 40 €/tonne. This shows good agreement with the results of the present techno-economic analysis, (33.5 €/tonne for Case 1). This is competitive with current state-of-the-art amine scrubbing carbon capture.
- FCs are commercially advantageous for power generation companies that are not tied to a particular type of fuel because FCs are flexible with regard to the feedstock. A company using SOFCs or MCFCs is prone to less risks of loss if one particular type of fuel becomes too expensive or even prohibited, since the fuel can be switched (with the appropriate processing and treatment steps).

8.3. Challenges

- Even though FCs show great potential to be a next generation CO_2 capture technology, there are still significant challenges that need to be addressed.
- As public policies begin to take shape, the concept of fuel cells combined with CO_2 capture could potentially become increasingly attractive since the FCs not only have higher electrical efficiencies in the power cycle, but also could potentially facilitate CO_2 capture at lower costs than conventional forms of fossil-fuel-based power generation. However, given their early stage of commercialisation, fuel cell systems will still require financial support mechanisms.
- An additional environmental benefit of FCs over conventional fossil power generation is the capability to minimise or even exclude the production of pollutants such as SO_X, NO_X and particulates. In moving to a clean economy the emphasis has been on CO_2 reduction, but given that every year almost 500,000 premature deaths have been linked to air pollution (European Environment Agency, 2018), there will potentially be a drive towards eliminating other air pollutants. This could potentially lead to the introduction of incentives / penalties that would contribute to an improved economic performance of FC systems.
- System manufacturing costs are expected to be reduced by increased mass manufacturing. The increasing average size of installations should fuel significant growth for the coming decade, but given price uncertainties, growth beyond that becomes less certain.
- Material and component costs are not expected to drop significantly with economies of scale. Further research is needed in that field in order to reduce FCs' installed costs and therefore make them competitive with other technology options.

9. RECOMMENDATIONS

- The results of this study show that FC with CCS hybrid cycles have the potential to be competitive with current state-of-the-art carbon capture technology but, as of yet there are no large-scale demonstrations which would be required to increase confidence in the technology and reduce the process risk. However, large projects are expected to be operational in the near future (Eisler, 2018) that will provide key performance data to enable further development of the next generation of carbon capture plants. This will increase confidence in the technology options and assist in overcoming the R&D challenges and barriers identified in the present study.
- The results presented here, referenced to common base cases, are intended to provide a guide towards the most promising configurations and technologies within the described limitations. It is recommended that the cases taken from literature are modelled to confirm performance as described, with bottom up estimates compiled for each case using a single source of data. This will ensure that the results are developed using the same basis and methodology and are therefore directly comparable.

REFERENCES

Adams, T. A. & Barton, P. I., 2010. High-efficiency power production from coal with carbon capture. *AIChE J.,* Volume 56.

Adams, T. A. & Barton, P. I., 2010. High-efficiency power production from natural gas with carbon capture. *J. Power Sources,* Volume 195.

Adams, T. A. & Barton, P. I., 2011. Combining coal gasification, natural gas reforming and solid oxide fuel cells for efficient polygeneration with CO2 capture and sequestration. *Fuel Processing Technology,* Issue 92, pp. 2105-2115.

Adams, T. A. & Nease J, 2014. Coal-fuelled systems for peaking power with 100% CO2 capture through integration of solid oxide fuel cells with compressed air energy storage. *Journal of Power Sources,* Volume 251, pp. 92-107.

Adams, T. A., Nease, J., Tucker, D. & Barton, P. I., 2013. Energy Conversion with Solid Oxide Fuel Cell Systems: A Review of Concepts and Outlooks for the Short- and Long-Term. *Ind. Eng. Chem. Res.,* Volume 52, pp. 3089-3111.

Aghaie, M., Mehrpooya, M. & Pourfayaz, F., 2016. Introducing an integrated chemical looping hydrogen production, inherent carbon capture and solid oxide fuel cell biomass fueled power plant process configuration. *Energy Conversion and Management,* Volume 124, pp. 141-154.

Anderson, G. L. & Garrigan, P. C., 1984. *Proceedings of the Symposium on Molten Carbonate Fuel.* Pennington, NJ, The Electrochemical Society, Inc..

Andresen, B., 2015. *BioZEG - demonstration of high efficiency carbon negative energy production.* Hannover, DE, Technical Forum, h2fc-fair.

Barelli, L., Bidini, G., Cinti, G. & Ottaviano, A., 2016. SOFC regulation at constant temperature: Experimental test and data regression study. *Energy Conversion and Management,* Volume 117, pp. 289-296.

Blamey, J., Anthony, E. J., Wang, J. & Fennell, P. S., 2010. The calcium looping cycle for large scale CO2 capture. *Progr Energy Combust,* 36(2), pp. 260-279.

Braun, R. J., 2010. Techno-economic optimal design of solid oxide fuel cell systems for micro-combined heat and power applications in the U.S. *J. Fuel Cell Sch. Tech.,* Volume 7.

Braun, R. J., Vincent, T. L., Zhu, H. & Kee, R. J., 2012. *Analysis, optimization, and control of solid oxide fuel cell systems.* New York: Elsevier.

Campanari, S., 2002. Carbon dioxide separation from high temperature fuel cell power plants. *J Power Sources,* 112(1), pp. 273-289.

Campanari, S., Chiesa, P. & Manzolini, G., 2010. CO2 capture from combined cycles integrated with molten carbonate fuel cells. *J. Greenh. Gas Control,* Volume 4, pp. 441-445.

Campanari, S., Chiesa, P., Manzolini, G. & Bedogni, S., 2014. Economic analysis of CO2 capture from natural gas combined cycles using Molten Carbonate Fuel Cells. *Applied energy,* Volume 130, pp. 562-573.

Campanari, S., Manzolini, G. & Chiesa, P., 2013. Using MCFC for high efficiency CO2 capture from natural gas combined cycles: Comparison of internal and external reforming. *Applied Energy,* Volume 112, pp. 772-783.

Campanari, S. et al., 2016. Predicting the ultimate potential of natural gas SOFC power cycles with CO2 capture e Part B: Applications. *Journal of Power Sources,* Volume 325, pp. 194-208.

Caprile, L., Passalacqua, B. & Torazza, A., 2011. Carbon capture: Energy wasting technologies or the MCFCs challenge?. *International Journal of Hydrogen Energy,* Issue 36, pp. 10269-10277.

Cheddie, D. F. & Murray, R., 2010. Thermo-economic modeling of a solid oxide fuel cell/gas turbine power plant with semi-direct coupling and anode recycling. *Int. J. Hydrogen Energy,* Volume 35, pp. 11208-11215.

Chiesa, P., Campanari, S. & Manzolini, G., 2011. CO2 cryogenic separation from combined cycles integrated with Molten Carbonate Fuel Cells. *Int. J. Hydrogen Energy,* 36(16), pp. 10355-10365.

Curletti, F. et al., 2015. Large size biogas-fed Solid Oxide Fuel Cell power plants with carbon dioxide management: Technical and economic optimization. *Journal of Power Sources,* Volume 294, pp. 669-690.

Curletti, F. et al., 2015. Large size biogas-fed Solid Oxide Fuel Cell power plants with carbon dioxide management: Technical and economic optimization. *Journal of Power Sources,* Volume 294, pp. 669-690.

Del Piano, G., 2012. *Tecnoeconomic analysis of the adoption of high temperature fuel cells for CO2 capture in large scale power plants,* Milan: Politecnico di Milano.

Diglio, G. et al., 2017. Techno-economic analysis of sorption-enhanced steam methane reforming in a fixed bed reactor network integrated with fuel cell. *Journal of Power Sources,* Issue 364, pp. 41-51.

Duan, L., Huang, K., Zhang, X. & Yang, Y., 2013. Comparison study on different SOFC hybrid systems with zero-CO2 emission. *Energy,* Volume 58, pp. 66-77.

Duan, L. et al., 2015. Study on a new IGCC (Integrated Gasification Combined Cycle) system with CO2 capture by integrating MCFC (Molten Carbonate Fuel Cell). *Energy,* Volume 87, pp. 490-503.

Duan, L. et al., 2016. Study on coal-fired power plant with CO2 capture by integrating molten carbonate fuel cell system. *Energy,* 117(2), pp. 578-589.

Duan, L. et al., 2016. Study on a novel pressurized MCFC hybrid system with CO2 capture. *Energy,* Volume 109, pp. 737-750.

Duan, L., Yue, L., Qu, W. & Yang, Y., 2015. Study on CO2 capture from molten carbonate fuel cell hybrid system integrated with oxygen ion transfer membrane. *Energy,* Volume 93, pp. 20-30.

Duan, L., Zhang, X. & Yang, Y., 2011. Exergy Analysis of a Novel SOFC Hybrid System with Zero-CO2 Emission. In: E. Benini, ed. *Advances in Gas Turbine Technology.* s.l.:IntechOpen, p. 81.

Dunbar, W. R., Lior, N. & Gaggioli, R. A., 1991. Combining fuel cells with fuel-fired power plants for improved exergy efficiency. *Energy,* 16(10), pp. 1259-1274.

Eisler, M., 2018. The carbon-eating fuel cell [Blueprints for a Miracle]. *IEEE Spectrum,* 55(6), pp. 22-76.

Element Energy & Vivid Economics, 2018. *Policy Mechanisms to support the largescale deployment of Carbon Capture and Storage (CCS),* s.l.: s.n.

European Environment Agency, 2018. *Air quality in Eurooe - 2018 report,* Luxemburg: Publications Office of the European Union.

Farooque, M., 1990. *Development on Internal Reforming Carbonate Fuel Cell Technology, Final report,* s.l.: U.S. DOE/METC, DOE/MC/23274-2941.

Farooque, M., 1991. *Development of Internal Reforming Carbonate Fuel Cell Stack Technology,* s.l.: US DoE/MC/23274-2941.

Fuel Cells and Hydrogen Joint Undertaking, 2015. *Advancing Europe's energy systems: Stationary fuel cells in distributed generation,* Luxemburg: Publications Office of the European Union.

Gazzani, M., Turi, D. M. & Manzolini, G., 2014. Techno-economic assessment of hydrogen selective membranes for CO2 capture in integrated gasification combined cycle. *Int. J. Greenh. Gas. Control,* Volume 20, pp. 293-309.

Gengo, T. et al., 2005. *Latest state of fuel cells that open the way to an environmentally friendly society in the 21 century,* s.l.: Mitsubishi Heavy Industries Tech. Rev..

Ghezel-Ayagh, H., 2017. *Electrochemical Membrane for Carbon Dioxide Capture and Power Generation,* s.l.: US DoE.

Ghezel-Ayagh, H., Jolly, S., Patel, D. & Steen, W., 2017. Electrochemical Membrane Technology for Carbon Dioxide Capture from Flue Gas. *Energy Procedia,* Volume 108, pp. 2-9.

Ghezel-Ayagh, H. et al., 2005. State of direct fuel cell/turbine systems development. *J. Power Sources,* Volume 152.

Greppi, P., Bosio, B. & Arato, E., 2013. Membranes and molten carbonate fuel cells to capture CO2 and increase energy production in natural gas power plants. *Industrial and engineering chemistry research,* Volume 52, pp. 8755-8764.

Grol, E., Keairns, D. & Newby, D., 2008. *Integrated gasification fuel cell (IGFC) system studies.* Pittsburgh, PA, 9th Annual SECA Workshop.

Hackett, G. A., Gerdes, K. & Chen, Y., 2015. Performance Impact associated with NI-Based SOFCs Fueled with Higher Hydrocarbon-Doped Coal Syngas. *Metallurgical and Materials Transactions E,* 2(1), pp. 58-69.

Hassanzadeh, H. & Mansouri, S. H., 2005. Efficiency of ideal fuel cell and Carnot cycle from a fundamental perspective. *J. Power and Energy,* Volume 219.

Henke, M. et al., 2012. Effect of pressure variation on power density and efficiency of solid oxide fuel cells. *Electochimica Acta,* Volume 66, pp. 158-163.

IEA, 2008. *CO2 capture and storage – a key carbon abatement option.* ISBN 978-92-64-041400 ed. s.l.:International Energy Agency.

IEAGHG, 2011. *The Costs of CO2 Storage: Post-demonstration CCS in the EU, July 2011*. Zero Emissions Platform (ZEP)

IEAGHG, 2018. *Effects of Plant Location on the Costs of CO2 Capture,* 2018-04, April 2018

IMF, 2018. *IMF Macroeconomic and Financial Data.* [Online]
Available at: http://data.imf.org/?sk=4C514D48-B6BA-49ED-8AB9-52B0C1A0179B&sId=1409151240976
[Accessed July 2018].

International Energy Agency (IEA), 2013. T*echnology Roadmap. Carbon capture and storage. 2013 edition,* s.l.: Paris: Organisation for Economic Co-operation and Development / International Energy Agency.

Jolly, S., Ghezel-Ayagh, H. & Stevenson, J. R., 2012. *Catalytic heat exchanger for SOFC balance-of-plant cost reduction.* Pittsburgh, PA, 13th Annual SECA Workshop.

Kee, R. J., Zhu, H., Sukeshini, A. M. & Jackson, G., 2008. Solid oxide fuel cells: operating principles, current challenges, and the role of syngas. *Combust. Sci. Technol,* Volume 180, pp. 1207-1244.

Klara, J. M., 2007. The potential of advanced technologies to reduce carbon capture costs in future IGCC power plants. *Energy Procedia,* Volume 1, p. 3827−3834.

Kneer, R. et al., 2010. OXYCOAL-AC: Towards an integrated coal-fired power plant process with ion transport membrane-based oxygen supply. *Eng. Environ. Sci.,* Volume 3, pp. 198-207.

Lanzini, A., Kreutz, T. G., Martelli, E. & Santarelli, M., 2014. Energy and economic performance of novel integrated gasifier fuel cell (IGFC) cycles with carbon capture. *International Journal of Greenhouse Gas Control,* Volume 26, pp. 169-184.

Lee, T. S., Chung, J. N. & Chen, Y., 2011. Design and optimization of a combined fuel reforming and solid oxide fuel cell system with anode off-gas recycling. *Energy Conv. Manage.,* Volume 52, pp. 3214-3226.

Leigh Fisher & Jacobs, 2016. *Electricity Generation Costs and and Hurdle Rates Lot 3: Non-Renewable Technologies,* s.l.: Department of Energy and Climate Change.

Liese, E., 2010. Comparison of preanode and postanode carbon dioxide separation for IGFC systems. *J. Eng. Gas Turbines Power,* 132(6), pp. 1-8.

Li, M., Brouwer, J., Rao, A. & Samuelsen, S., 2011. Application of a detailed dimensional solid oxide fuel cell model in integrated gasification fuel cell system design and analysis. *Journal of Power Sources,* 196(14), pp. 5903-5912.

Li, M., Rao, A. D. & Samuelsen, G. S., 2012. Performance and costs of advanced sustainable central power plants with CCS and H2 co-production. *Appl Energy,* 91(1), pp. 43-50.

Lim, T. H. et al., 2008. Operating characteristics of a 5kW class anode-supported planar SOFC stack for a fuel cell/gas turbine hybrid system. *Int. J. Hydrogen Energy,* Volume 33, pp. 1076-1083.

Lu, X., 2017. *Allam Cycle Coal - A New Clean Coal Power Cycle,* s.l.: U.S. - China Clean Coal Industry Forum.

Lygre, A., Ce, M., Vik, A. & Byrknes, J., 2001. *Solid oxide fuel cell power plants with integrated CO2 capture.* Gothenburg, Sweden, Second Nordic minisymposium on carbon dioxide capture and storage.

Madi, H. et al., 2015. Solid oxide fuel cell anode degradation by the effect of siloxanes. *Journal of Power Sources,* Issue 279, pp. 460-471.

Mamaghani, A. H., Najaf, B., Shirazi, A. & Rinaldi, F., 2015. Exergetic, economic, and environmental evaluations and multi-objective optimization of a combined molten carbonate fuel cell-gas turbine system. *Appl. Therm. Eng.,* Volume 77, pp. 1-11.

Manzolini, G. et al., 2012. CO2 separation from combined cycles using molten carbonate fuel cells. *J. Fuel Cell Sci. Technol.,* Volume 9, pp. 1-8.

Marquez, A. I. et al., 2007. Effects of coal syngas and H2S on the performance of solid oxide fuel cells: part 2. Stack tests. *J. Power Sources,* Volume 164, pp. 659-667.

Maurstad, O. et al., 2005. SOFC and gas turbine power systems- evaluation of configurations for CO2 capture. *Greenhouse Gas Control Technologies,* Volume 1, pp. 273-281.

Milewski, J., Lewandowski, J. & Miller, A., 2009. Reducing CO2 emissions from a coal fired power plant by using a molten carbonate fuel cell. *Chem Process Eng.,* Volume 30, pp. 341-350.

Morita, H., Kawase, M., Mugikura, Y. & Asano, K., 2010. Degradation mechanism of molten carbonate fuel cell based on long-term performance: Long-term operation by using bench-scale cell and post-test analysis of the cell. *Journal of power sources,* Volume 195, pp. 6988-6996.

Morita, H., Kawase, M., Mugikura, Y. & Asano, K., 2010. Degradation mechanism of molten carbonate fuel cell based on long-term performance: Long-term operation by using bench-scale cell and post-test analysis of the cell. *Journal of Power Sources,* Volume 195, p. 6988–6996.

NETL, 2009. *Integrated gasification fuel cell performance and cost assessment,* s.l.: DOE/NETL.

NETL, 2015. *Cost and Performance Baseline for Fossil Energy Plants Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity Revision 3,* s.l.: DOE/NETL.

NETL, 2015. *Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b - Year Dollar Update,* s.l.: DOE/NETL.

Nichols, D. K., Agnew, G. & Strickland, D., 2008. *Outlook and application status of the Rolls-Royce fuel cell systems SOF.* Pittsburgh, PA, s.n., pp. 1-3.

NREL, 2010. *Molten Carbonate and Phosphoric Acid Stationary Fuel Cells: Overview and Gap Analysis,* Golden: National Renewable Energy Laboratory.

OFX, 2018. *OFX Historical Exchange Rates.* [Online]
Available at: https://www.ofx.com/en-gb/forex-news/historical-exchange-rates/
[Accessed July 2018].

Pascenti, M., Ferrari, M. L., Magistri, L. & Massardo, A. F., 2007. *Micro gas turbine based test rig for hybrid system emulation.* Montréal, Canada, Proceedings ASME Turbo Expo.

Pilatowsky, I., Romero, R. J., Isaza, C. A. & Gamboa, S. A., 2011. *Cogeneration Fuel Cell-Sorption Air Conditioning Systems.* s.l.:Springer.

Restrepo, B., Banta, L. & Tucker, D., 2011. *Characterization of a solid oxide fuel cell gas turbine hybrid system based on a factorial design of experiments using hardware simulation.* Washington, DC, Proceedings of the ASME 2011 5th Annual Energy Sustainability Conference and 9th Annual Fuel Cell Science and Engineering Technology Conference.

Rexed, I. et al., 2015. Molten carbonate fuel cells for CO2 separation and segregation by retrofitting existing plants - An analysis of feasible operating windows and first experimental findings. *International journal of greenhouse gas control,* Volume 35, pp. 120-130.

Riensche, E. et al., 2000. Clean combined-cycle SOFC power plant − cell modelling and process analysis. *J. Power Sources,* Volume 86, p. 404.

Riensche, E., Meusinger, J., Stimming, U. & Unverzagt, G., 1998. Optimization of a 200 kW SOFC cogeneration power plant. Part II: variation of the flowsheet. *J. Power Sources,* Volume 71, pp. 306-314.

Romano, M., Spallina, V. & Campanari, S., 2011. Integrating IT-SOFC and gasification combined cycle with methanation reactor and hydrogen firing for near zero-emission power generation from coal. *Energy Procedia,* Volume 4, pp. 1168-1175.

Samanta, S. & Ghosh, S., 2016. A thermo-economic analysis of repowering of a 250MW coal fired power plant through integration of Molten Carbonate Fuel Cell with carbon capture. *International journal of greenhouse gas control,* Volume 51, pp. 48-55.

Samanta, S. & Ghosh, S., 2017. Techno-economic assessment of a repowering scheme for a coal fired power plant through upstream integration of SOFC and downstream integration of MCFC. *International Journal of Greenhouse Gas Control,* Volume 64, pp. 234-245.

Sanchez, D., Monje, B., Chacartegui, R. & Campanari, S., 2013. Potential of molten carbonate fuel cells to enhance the performance of CHP plants in sewage treatment facilities. *International journal of hydrogen energy ,* Volume 38, pp. 394-405.

Santarelli, M. et al., 2013. SOFCOM Project: Proof-of-Concept of WWTU Plant Feeding a SOFC CHP System Integrated with CO2 Removal. *ECS Transactions,* 57(1), pp. 185-195.

Scataglini, R. et al., 2015. *A Total Cost of Ownership Model for Solid Oxide Fuel Cells in Combined Heat and Power and Power-Only Applications,* s.l.: Ernest Orlando Lawrence Berkeley National Laboratory.

Shelley, S., 2009. Oxygen and nitrogen: onward and upward. *Chem. Eng. Prog.,* 105(1).

Siefert, N. & Litster, S., 2013. Exergy and economic analyses of advanced IGCC-CCS and IGFC-CCS power plants. *Applied Energy,* Volume 107, pp. 315-328.

Spallina, V., Romano, M., Campanari, S. & Lozza, G., 2011. A SOFC-based integrated gasification fuel cell cycle with CO2 capture. *J. Eng. Gas Turbines Power,* Volume 133.

Spinelli, M. et al., 2018. Molten Carbonate Fuel Cells for Retrofitting Postcombustion CO2 Capture in Coal and Natural Gas Power Plants. *Journal of Electrochemical Energy Conversion and Storage,* Volume 15, pp. 031001-031001-15.

SRI International, 2008. *Effect of Coal Contaminants on Solid Oxide Fuel System Performance and Service Life ,* Morgantown: US Department of Energy.

Sugiura, K., Takei, K., Tanimoto, K. & Miyazaki, Y., 2003. The carbon dioxide concentrator by using MCFC. *Journal of Power Sources,* Volume 118, pp. 218-227.

Thijssen, J., 2011. *Solid oxide fuel cells and critical materials: A review of implications,* s.l.: US Department of Energy/NETL.

Thijssen, J. H., 2009. *Stack Operating Strategies for Central Station SOFC.* s.l., Fuel Cell seminar, NETL, US DOE.

Traverso, A., Tucker, D. & Haynes, C., 2011. *Preliminary experimental results of IGFC operation using hardware simulation.* Vancouver, Canada, ASME/IGTI Turbo Expo.

Trembly, J. P., Gemmen, R. S. & Bayless, D. J., 2007. The effect of coal syngas containing HCl on the performance of solid oxide fuel cells: investigations into the effect of operational temperature and HCl concentration.. *J. Power Sources,* Volume 169, pp. 347-354.

Udengaard, N. R., Olsen, A. & Wix-Nielsen, C., 2006. *High temperature methanation process.* Pittsburgh, s.n.

US DOE Fuel Cell Technologies, 2015. *DOE Technical Targets for Fuel Cell Systems for Stationary (Combined Heat and Power) Applications.* [Online]
Available at: https://www.energy.gov/eere/fuelcells/doe-technical-targets-fuel-cell-systems-

stationary-combined-heat-and-power
[Accessed 15 November 2018].

US DoE, 2004. *Fuel Cell Handbook (Seventh Edition).* s.l.:EG&G Technical Services, Inc..

Verma, A., Rao, A. D. & Samuelsen, G. S., 2006. Sensitivity analysis of a Vision 21 coal based zero emission power plant. *J Power Sources,* Volume 158, pp. 417-427.

Wang, J., 2015. Barriers of scaling-up fuel cells: Cost, durability and reliability. *Energy,* Volume 80, p. 509e521.

Woods, M. C. et al., 2012. Cost and Performance Baseline for Fossil Energy Plants. Volume 1: Bituminous Coal and Natural Gas to Electricity Final Report. *DOE/NETL-2007/1281, Revision 1. US Energy Information Administration..*

Wuilleim, Z., 2009. *Experimental and Modeling Investigations on Local Performance and Local Degradation in Solid Oxide Fuel Cells.* Master Thesis: EPFL.

Zhang, X. et al., 2010. A review of integration strategies for solid oxide fuel cells. *J Power Sources,* Volume 195.

Ziock, H. et al., n.d. *Zero Emission Coal, A New Approach and Why It Is Needed.* s.l., s.n.

GLOSSARY OF TERMS

AB	After-burner
AFC	Alkaline Fuel Cell
APH	Air Preheater
ASR	Overall Stack Resistance
ASU	Air Separation Unit
AT	Air Turbine
BEC	Bare Erected Cost
CCS	Carbon Capture & Storage
CEPACS	Combined Electric Power and Carbon-dioxide Separation
CLC	Chemical Looping Combustion
COE	Cost of Electricity
DBFB	Dual Bubbling Fluidised Bed
DIR	Direct Internal Reforming
ECM	Electrochemical Membrane
EOR	Enhanced Oil Recovery
EPC	Engineering, Procurement and Construction
ETS	EU Emissions Trading System
FCs	Fuel Cells
FP	Fuel Preheater
GEE	General Electric Energy
GHGs	Greenhouse Gases
GSM	Gas Separation Membrane
GT	Gas Turbine
HRSG	Heat Recovery Steam Generator
HHV	High Heating Value
HX	Heat Exchanger
IGCC	Integrated Gasification Combined Cycle
IGFC	Integrated Coal Gasification Fuel Cell
IIR	Indirect Internal Reforming
ITM	Ion Transfer Membrane
kW	kilowatt
LCOE	Levelised Cost of Electricity
LHV	Low Heating Value
MCFC	Molten Carbonate Fuel Cell
MEA	monoethanolamine
MW	Megawatt
NETL	National Energy Technology Laboratory
NGCC	Natural Gas Combined Cycle
NGFC	Natural Gas Combined Cycle Fuel Cell
NPV	Net Present Value
OC	Open Cycle
OCM	Oxygen-Conducting Membrane
PAFC	Phosphoric Acid Fuel Cell
PC	Pulverised Coal
PEMFC	Proton Exchange Membrane Fuel Cell
SC	Supercritical
SER	Sorption Enhanced Reforming
SE-SMR	Sorption Enhanced Steam Methane Reforming
SOFC	Solid Oxide Fuel Cell

SPECCA	Specific primary energy consumption for CO_2 avoided
T&EFS	Technical and Economic Feasibility Study
TOC	Total Operational Cost
TPC	Total Plant Cost
TRL	Technology Readiness Level
U_{Air}	Air Utilisation
U_{CO2}	CO_2 Utilisation
U_f	Fuel Utilisation
US DoE	U. S. Department of Energy
WGS	Water-Gas-Shift
ZEC	Zero-Emission Coal
ZEG	Zero Emission Gas

Appendix 1

Review of MCFC and SOFC technologies with CO_2 capture

Table A: Literature study of SOFC + CCS

	Process Description						SOFC Operational data					Economic performance		Reference	
No.	**Fuel**	**Details**	**type of gasifier**	**(Syn)gas cleanup/Reforming**	**CO2 Capture Technology**	**WGS?**	**Voltage (V)**	**Uf**	**T (°C)**	**P (bar)**	**SOFC capacity (kW)**	**Efficiency (HHV)**	**LCOE ($/MWh)**	**Refs**	**Year**
SO1	Coal	SOFC+GT	Oxygen-blown, fluidized bed	No Data	WGS followed by membrane	After SOFC	No Data	35.60%	1000	15	110000	46.30%	No Data	Kuchonthara et al.	2005
SO2	Coal	SOFC+GT	Oxygen-blown, fluidized bed	Warm gas cleaning	WGS followed by H2 membrane	After SOFC	0.75	85%	800	18.8	260000	50.30%	No Data	Verma et al.	2006
SO3	Coal	SOFC+GT	Catalytic hydrogasifier	Warm humid gas cleaning	Oxycombustion followed by condensation	-	0.8	82%	650	1.8	239150	49.40%	88	NETL Ashok et al.	2009 2010
SO4	Coal	SOFC+GT	Catalytic hydrogasifier	Warm humid gas cleaning	Oxycombustion followed by condensation	-	0.87	85	650	18.9	221060	56.20%	79	NETL Ashok et al.	2009 2010
SO5	Coal	SOFC+GT	Catalytic hydrogasifier	Low temperature gas cleaning	MDEA stripper (90%)	After SOFC	0.8	75	850	10	230000-250000	57-61.5%	No Data	Ashok Li et al. Li et al.	2010 2011 2011
SO6	Coal	SOFC+GT	Conoco Philips E-Gas gasifier	Low temperature gas cleaning	Condenser after anode outlet	-	0.75	80%	650-800	3.8	621000	43-47%	No Data	Liese	2010
SO7	Coal	SOFC+GT	Conoco Philips E-Gas gasifier	Low temperature gas cleaning	Selexol before SOFC	Before SOFC	0.75	80%	650-800	9	627000	52%	No Data	Liese	2010
SO8	Coal	SOFC+GT	Conoco Philips E-Gas gasifier	Low temperature gas cleaning	Selexol before SOFC	-	0.75	80%	650-800	4	753000	53%	No Data	Liese	2010
SO9	Coal	SOFC+GT	Oxygen-blown entrained	No Data	Condenser after anode outlet	-	0.78	89.20%	800	20	310000-330000	45–46%	No Data	Spallina et al.	2011
SO10	Coal	SOFC+GT	Oxygen-blown entrained	No Data	Oxycombustion followed by condensation	-	0.68-0.79	66-86%	775	3-5	111000	45–48%		Braun et al.	2012
SO11	Coal	SOFC+GT	GE Radiant- Only gasifier	Low temperature gas cleaning	Oxidation unit followed by flash cascade	Before SOFC	0.69	99.50%	1000	20	563000	45%	63.5	Adams and Barton Adams and Barton	2010 2011
SO12	Coal	SOFC+GT	Catalytic gasifier	Low temperature gas cleaning	Oxycombustion followed by condensation	-	No Data	No Data	No Data	No Data	550000 net plant	55%	No Data	Grol et al.	2008
SO13	Coal	SOFC	Catalytic gasifier	Warm gas cleaning	Oxycombustion followed by condensation	-	No Access	No Access	No Access	No Access	632000	56-59%	No Access	Ghezel-Ayagh et al. Gray et al.	2011 2011 2010
SO14	Coal	SOFC	Non-catalytic gasifier	No Data	Selexol before SOFC	Before SOFC	No Data	No Data	No Data	No Data	No Data	43%	No Data	Grol et al.	2008
SO15	Coal	SOFC	Non-catalytic gasifier	No Data	Oxycombustion followed by condensation	-	No Data	No Data	No Data	No Data	No Data	44%	No Data	Grol et al.	2008
SO16	Coal	SOFC	No Data	Low temperature gas cleaning	Pre-combustion	-	0.773	70%	900	19	290000	40%	No Data	Park et al.	2011

SO17	Coal	SOFC	No Data	Low temperature gas cleaning	Oxycombustion followed by condensation	-	0.773	70%	900	19	238000	47%	No Data	Park et al.	2011
SO18	Coal	SOFC	Entrained-flow shell	Low temperature gas cleaning	Pre- and post- SOFC absorption	Before SOFC	0.812	74%	800	34.15	306000	46%	No Data	Romano et al.	2009
SO19	NG	SOFC+GT	-	-	absorption after GT/SOFC	-	No Data	No Data	No Data	No Data	2217	57-58%	No Data	Duan et al.	2011
SO20	NG	SOFC+GT	-	-	Single condenser after GT/ SOFC	-	No Data	No Data	No Data	No Data	1500 (net)	53%	No Data	Massardo et al. Franzoni et al.	2000 2008
SO21	NG	SOFC+GT	-	Steam	Absorption after SOFC/GT	-	No Data	No Data	No Data	No Data	No Data	No Data	No Data	Varatharajan et al.	2005
SO22	NG	SOFC+GT	-	ATR	Absorption after SOFC/GT	Before turbine	No Data	No Data	No Data	No Data	1500	44.70%	No Data	Massardo et al. Franzoni et al.	2000 2008
SO23	NG	SOFC	-	-	Absorption after SOFC	After SOFC	0.64	85%	1000	1.3	54000	63%	No Data	Campanari	2002
SO24	NG	SOFC	-	-	Heat exchange after SOFC	-	0.705	85%	1000	1.35	5000-54000	61-72%	No Data	Kuramochi et al. Campanari	2011 2002
SO25	NG	SOFC	-	Steam	Single condenser after SOFC	-	0.7	85%	900	5	54	44-58%	No Data	Riensche et al.	2000
SO26	NG	SOFC	-	Steam	Single condenser after SOFC	-	No Data	No Data	No Data	No Data	No Data	No Data	No Data	Galloway et al.	2007
SO27	NG	SOFC	-	POx, ATR	Multistage flash after SOFC	Before SOFC	0.69	92%	950	10.1	562000-614000	61-67%	No Data	Adams et al.	2010
SO28	NG	SOFC	-	Steam	Multistage flash after SOFC	Before SOFC	0.69	92%	950	10.1	752000	73-74%	64.5	Adams et al.	2010
SO29	NG	SOFC, After burner, HRSG, MCFC	-	Direct Internal Reforming (DIR)	MCFC Cathode		SOFC: 0.8 MCFC: 0.7	SOFC: 85% MCFC: 75%	SOFC: 800 MCFC: 650	SOFC: atm MCFC: atm	175000 (net)	46.70%		Samanta, Gosh	2017
SO30	Co-firing (70% woodpellets, 30% coal)	IGCC SOFC retrofitted, with SOFC not the main power unit	Shell Coal Gasification Process (SCGP)	Syngas Cooler (SGC)	Partial oxy-combustion	Before SOFC	0.83	85%	900	No Data	47550	40.77%	No Data	Thattai, et al.	2017
SO31	Co-firing (70% woodpellets, 30% coal)	IGFC pressurised SOFC retrofitted, with SOFC main power unit	Shell Coal Gasification Process (SCGP)	Syngas Cooler (SGC)	Full oxy-combustion	Before SOFC	0.83	85%	900	No Data	47550	47.96%	No Data	Thattai, et al.	2017
SO32	NG	Ambient pressure SOFC + steam based Rankine cycle	-	Adiabatic pre-reforming & DIR	Cryogenic CO2 condensation	After SOFC	0.86	85%	735	1.2	68060	64.59%	No Data	Campanari, et al.	2016
SO33	NG	Pressurised SOFC + Brayton cycle	-	Adiabatic pre-reforming & DIR	Cryogenic CO2 condensation	After SOFC	0.86	85%	735	20	68060	63.99%	No Data	Campanari, et al.	2016

SO34	Coal	SOFC + GT + ST	Commercial, pressurized, O2-blown, entrained flow, dry-feed Shell gasifier	AGR --> Internal reforming	Post-anode oxy-combustion	-	?	85%	800	20	702800	51.40%	78.8	Lanzini, et al	2014
SO35	Coal	SOFC + GT + ST	Commercial, pressurized, O2-blown, entrained flow, dry-feed Shell gasifier	WGS --> AGR --> Methanator --> Internal reforming	Post-anode oxy-combustion	Before SOFC	?	85%	800	20	706100	47.20%	89.4	Lanzini, et al	2014
SO36	Coal	SOFC + GT + ST	Commercial, pressurized, O2-blown, entrained flow, dry-feed Shell gasifier	AGR --> Methanator --> Internal reforming	Post-anode oxy-combustion	-	?	85%	800	20	693100	52.10%	77.6	Lanzini, et al	2014
SO37	Biogas	SOFC	-	Integrated external reformer + Internal reforming (50%)	Post-anode oxy-combustion --> Condensation --> Non-porous membrane	-	0.8	75%	800	atm	1000	48.36%	NPV (M$): 0.6001	Curletti et al	2015
SO38	Biogas	SOFC	-	Integrated external reformer + Internal reforming (50%)	Post-anode oxy-combustion --> Condensation --> Non-porous membrane	-	0.8	75%	800	2	1000	49.24%	NPV (M$): 1.6335	Curletti et al	2015
SO39	Biogas	SOFC	-	Integrated external reformer + Internal reforming (50%)	Post-anode oxy-combustion --> Condensation --> Non-porous membrane	-	0.8	75%	800	6	1000	48.85%	NPV (M$): 1.8551	Curletti et al	2015
SO40	Coal	SOFC + ST	Dry-fed type gasifier	Internal reforming	Chemical Looping (NiO carrier)	Internal	No Data	85%	900	15	7652.4	49.80%	No Data	Chen et al.	2015
SO41	Coal	SOFC + ST	Dry-fed type gasifier	Internal reforming	Chemical Looping (Fe2O3 carrier)	Internal	No Data	85%	900	15	7652.4	49.20%	No Data	Chen et al.	2015
SO42	Coal	SOFC + ST	Dry-fed type gasifier	Internal reforming	Chemical Looping (CuO carrier)	Internal	No Data	85%	900	15	7652.4	47.60%	No Data	Chen et al	2015
SO43	NG	SOFC+GT	-	Internal Prereformer	2nd SOFC afterburner + h2o condensation	Internal	0.7	85%	1000	9.03	18310	67.40%	No Data	Maurstad et al.	2005
SO44	NG	SOFC+GT	-	Internal Prereformer	Oxygen separation membrane reactor afterburner + h2o condensation	Internal	0.7	85%	1000	9.03	17080	64.80%	No Data	Maurstad et al.	2005
SO45	NG	SOFC+GT	-	Internal Prereformer	Hydrogen separation membrane reactor afterburner + H20 condensation	Internal	0.7	85%	1000	9.03	17080	64.80%	No Data	Maurstad et al.	2005
SO46	NG	SOFC + GT	-	External pre-reformer	Oxyfuel with steam injection	-	0.635	85%	910	3	141960	62.07%	No Data	Duan et al.	2013
SO47	NG	SOFC + GT	-	External pre-reformer	Oxyfuel with CO2 injection	-	0.635	85%	910	3	141960	63.32%	No Data	Duan et al.	2013
SO48	NG	SOFC + GT	-	External pre-reformer	Oxyfuel with heat exchange layout	-	0.635	85%	910	3	141960	63.10%	No Data	Duan et al.	2013
SO49	Coal	SOFC + Brayton cycle	Exxon single-stage, fluidised bed catalytic gasifier	Internal reforming	Pre-anode capture using MgO and CaO	Internal	0.7	70% (single pass)	850	5	500000	58%	52 ± 17 (2007$, EOR sequestration)	Siefert et al.	2013
SO50	Coal	SOFC + Brayton cycle	Exxon single-stage, fluidised bed catalytic gasifier	Internal reforming	Pre-anode capture using MgO and CaO	Internal	0.7	70% (single pass)	850	5	500000	58%	60 ± 17 (2007$, Saline sequestration)	Siefert et al.	2013

SO51	Coal	SOFC + Brayton cycle	CaO-looping moving bed gasifier	Internal reforming	Pre-anode capture using CaO (75% capture)	Internal	0.7	80%	885	3	116000	60.20%	No Data	Siefert et al.	2014
SO52	Coal	SOFC + ST	Coal gasification and iron oxide reactions using a fluidised or moving bed (no gaseous oxygen used)	Internal reforming	Pre-anode Chemical looping Hydrogen generation + post-anode CO2 HRSG	-	0.7	80%	900	3.5	14610	41.59%	No Data	Chen et al.	2017
SO53	NG	SOFC + GT + ST	-	No data	Oxyfuel electrochemical afterburner (100% capture)	No Data	0.65	85%	No Data	2.5	20000	64%	59.2	Lokurlu et al.	2005
SO54	NG	SOFC + GT + ST	-	No data	Oxyfuel electrochemical afterburner - Additional afterburner (80% capture)	No Data	0.65	85%	No Data	2.5	20000	59%	51.8	Lokurlu et al.	2005
SO55	NG	SOFC + GT + ST	-	No data	Oxyfuel electrochemical afterburner - Mixed oxide conductor (MOC)	No Data	0.65	85%	No Data	2.5	20000	60%	60.68	Lokurlu et al.	2005
SO56	NG	SOFC + GT + ST	-	No data	Oxyfuel electrochemical afterburner - Oxygen pump	No Data	0.65	85%	No Data	2.5	20000	53%	65.12	Lokurlu et al.	2005
SO57	NG	SOFC + GT + TRCC	-	Internal reforming	Transcritical Co2 Cycle (TRCC)	Internal	0.69	85%	831	No Data	1754	69.26%	No Data	Meng et al.	2017
SO58	Coal	SOFC + GT	Shell commercial dry fed gasifier	Internal reforming	Pre-anode Chemical looping (Iron oxide) Hydrogen generation + post-anode CO2 HRSG	-	0.7	75%	900	20	3666.31	43.53%	No Data	Chen et al.	2012
SO59	Biomass	SOFC + GT + ST	Steam gasification	External pre-reformer	Pre and post-anode Chemical Looping using CaO for H2 production.	Before SOFC	0.72	85%	850	12	6050	55.80%	No Data	Aghaie et el.	2016

Table B: Literature study of MCFC + CCS

	Process Description					MCFC Operational data						Economic performance				Reference	
No.	**Fuel**	**Details**	**Cleanup/Reforming**	**WGS**	**CO2 Capture Technology**	**Voltage (V)**	**Uf**	**UCO2**	**T (°C)**	**P (bar)**	**MCFC capacity (kW)**	**Efficiency (HHV)**	**LCOE ($/MWh)**	**Cost of CO2 avoided ($/tonCO2)**	**CCR**	**Refs**	**Year**
MC1	NG	MCFC placed between GT and HRSG	Desulfurisation (ZnO absorption beds or active carbon filtering) Internal reforming	Internal	Cryogenic CO2 separation	0.709	75.00%	75.00%	650	No data	94800	58.88%	84.66 (EUR)	120.4 (EUR)	71.90%	Campanari et al.	2014
MC2	NG	MCFC placed between GT and HRSG	Desulfurisation (ZnO absorption beds or active carbon filtering) Internal reforming	Internal	Oxy-combusiton	0.702	75.00%	75.00%	650	No data	78800	No data	81.53 (EUR)	107.7 (EUR)	72.50%	Campanari et al.	2014
MC3	NG	MCFC placed between GT and HRSG	Desulfurisation (ZnO absorption beds or active carbon filtering) External reforming thermally integrated with the power cycle (double fuel-cell layout)	External	Cryogenic CO2 separation	Stack 1: 0.75 Stack 2: 0.70	75.00%	75.00%	650	No data	Stack 1: 86900 Stack 2: 32900	58.63%	No data	No data	76.32%	Campanari et al.	2013
MC4	NG	MCFC placed between GT and HRSG	Desulfurisation (ZnO absorption beds or active carbon filtering) External reforming thermally integrated with the power cycle (double fuel-cell layout)	External	Oxy-combusiton	Stack 1: 0.75 Stack 2: 0.70	75.00%	75.00%	650	No data	Stack 1: 39000 Stack 2: 41800	57.53%	No data	No data	Stack 1: 72.6% Stack 2: 75.5%	Campanari et al. Brenna et al.	2013
MC5	NG	MCFC + Afterburner + HRSG ST	Pre-reformer	External	Cryogenic air separation	0.66	85.00%	85.00%	650	No data	86700	60.94%	No data	No data	85%	Duan et al.	2015
MC6	NG	MCFC + Afterburner + HRSG ST	Pre-reformer	External	Oxygen Ion Transfer Membrane	0.66	85.00%	85.00%	650	No data	86700	62.68%	No data	No data	85%	Duan et al.	2015
MC7	Anode: NG Cathode: Exhaust of coal fired pp	Steam cycle + MCFC + HRSG ST	Cathode: fed with the existing boiler exhaust after desulphurisation, dilution with fresh air and pre-heating External reformer	External	Oxy-combustion + moisture separation	0.7	75%	52.06%	650	No data	100300	36.76%	81.74 (COE)	43.96	70%	Samanta & Ghosh	2016
MC8	Anode: NG Cathode: Exhaust of IGCC	MCFC placed between GT and HRSG	Pre-reformer	External	Afterburner + Oxycombustion	No data	80%	85%	650	1	330200	47.31%	103.85 (COE)	63.73	No data	Duan et al.	2015
MC9	NG	MCFC placed between GT and HRSG	Internal reforming	Internal	Condensation	0.656	No data	No data	No data	No data	83247	No data	No data	No data	No data	Yazdanfar et al.	2015
MC10	Anode: NG Cathode: Exhaust of coal fired pp	Steam cycle + MCFC	Anode: Desulphuriser Cathode: Gas treatment (SO2, ash) Internal reforming	Internal	Physical Absorption	0.76	78.50%	76%	650	1.013	198000	45.77%	No data	No data	76.90%	Campanari	2002
MC11	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp	No data	No data	No data	0.5-0.75	20-100%	20-100%	650	1	94500	No data	No data	No data	47.50%	Sugiura et al.	2003
MC12	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp	No data	No data	No data	0.5-0.75	20-100%	20-100%	650	1	94500	No data	No data	No data	90.00%	Sugiura et al.	2003

MC13	NG	MCFC placed after GT	External reformer	External	Condensation	No data	47% 75%	<55%	635	1	2000	No data	No data	No data	53.00%	Lusardi et al.	2004
MC14	NG	MCFC placed after GT	External reformer	External	Condensation	0.8	No data	No data	No data	No data	1600	No data	No data	No data	50.00%	Amorelli et al.	2004
MC15	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp	Internal reforming	Internal	Condensation	No data	No data	No data	No data	No data	No data	40.00%	No data	No data	39.00%	Milewski et al.	2009
MC16	NG	MCFC placed between GT and HRSG	Desulfurisation (ZnO absorption beds or active carbon filtering) Internal reforming	Internal	Condensation	0.699	75%	75%	650	No data	88600	58.00%	No data	No data	80.00%	Campanari et al.	2010
MC17	NG	MCFC placed between GT and HRSG	External reformer	External	Membrane	No data	75%	60%	690	No data	212000	52.50%	COE: 70-72.3 (EUR)	45.3 (EUR)	58.10%	Greppi et al.	2013
MC18	HC	MCFC placed after gas turbine	Internal reforming	Internal	Compression	No data	75%	75%	650	1	1800	41.30%	No data	No data	76.00%	Sanchez et al.	2014
MC19	No data	No data	No data	No data	Anode outlet gas (two different streams) burnt with air - Condensation	No data	No data	No data	No data	No data	500	52.00%	No data	No data	No data	De Lorenzo et al.	2010
MC20	Coal	Absorption Enhanced Reforming (AER) + MCFC	ZnO desulphurisation External reformer	External	Ca looping	No data	95%	85%	650	1.04	200	44.70%	Specific investment (£/kWe): 11642	No data	No data	Wang et al.	2006
MC21	NG	MCFC placed between GT and HRSG ST	External reformer	External	Condensation	No data	No data	No data	No data	No data	2400	48.20%	No data	No data	583 t / y	Pak et al.	2009
MC22	Coal	MCFC + Rankine cycle	Conventional desulfurisation Internal reforming	Internal	Cryogenic CO2 capture	0.784	80%	80%	650	4.5	6000	45% (net power) 71% (with heat recovery)	No data	No data	No data	Mehrpooya et al.	2017
MC23	NG	MCFC placed between GT and HRSG	Pre-reformer	External	Condensation	No data	No data	No data	650	No data	50500	55.61%	No data	No data	45%	Duan et al.	2014
MC24	NG	MCFC placed between GT and HRSG	Pre-reformer	External	Condensation	No data	No data	No data	650	No data	50500	54.96	101.9 (COE)	108.4	85%	Duan et al.	2014
MC25	NG	MCFC + Afterburner + HRSG ST	Pre-reformer	External	Oxyfuel + condensation	0.763	85%	85%	650	6	106420	68.43%	164.3 (COE)	No data	90%	Duan et al.	2016
MC26	Anode: NG Cathode: Exhaust of coal fired pp	Coal fired pp + MCFC + Afterburner + HRSG ST	Pre-reformer	External	Oxyfuel + condensation	0.68	85%	85%	650	1	313890	45.62%	82.556 (COE)	60.722	85%	Duan et al.	2016

MC27	NG	MCFC placed after gas turbine	External reformer	External	Membrane	0.6-0.8	No data	55.70%	No data	Pressurised stack	135000	48.60%	No data	No data	67%	Caprile et al.	2011
MC28	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp	External reformer	External	Membrane	0.6-0.8	No data	55.70%	No data	Pressurised stack	172000	38.10%	No data	No data	68.90%	Caprile et al.	2011
MC29	NG	MCFC placed after gas turbine	External reformer	External	Condensation	0.706	70.45	405	695	1	386	44.65	No data	No data	61.90%	Desideri et al.	2012
MC30	NG	MCFC placed after gas (biogas) turbine	External reformer	External	Oxy-combusiton	0.636	70%	80%	650	2.2	1063	31%	CAPEX: $3.19M	No data	89%	Barelli et al.	2016
MC31	Biogas	MCFC placed after gas (biogas) turbine	External reformer	External	Oxy-combusiton	0.574	77%	80%	650	2.2	1049	31%	CAPEX: $3.60M	No data	86.50%	Barelli et al.	2016
MC32	Bio-optimised gas	MCFC placed after gas (biogas) turbine	External reformer	External	Oxy-combusiton	0.592	75%	80%	650	2.2	1068	29.50%	CAPEX: $3.65M	No data	95.10%	Barelli et al.	2016
MC33	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp (Maximum efficiency case)	Internal reforming	Internal	Condensation	0.5	86%	No data	648	No data	No data	41%	No data	No data	52%	Milewski & Lewandowski	2012
MC34	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp (Maximum CO2 reduction case)	Internal reforming	Internal	Condensation	0.4	90%	No data	643	No data	No data	39%	No data	No data	74%	Milewski & Lewandowski	2013
MC35	Anode: NG Cathode: Exhaust of coal fired pp	MCFC using the exhaust of a coal fired pp (Optimal case)	Internal reforming	Internal	Condensation	0.44	88%	No data	650	No data	No data	37%	No data	No data	70%	Milewski & Lewandowski	2014

Appendix 2

FC Possible Configurations and their Dependencies

A	Steam reforming	Adiabatic pre-reforming	Excludes option B	✓ Converts higher hydrocarbons and reduces the risk of carbon deposition in the FC or in the downstream reformer.
B		Heat exchange pre-reforming by waste heat recovery	Excludes options A	✓ Converts higher hydrocarbons and reduces the risk of carbon deposition in the FC or in the downstream reformer. ✓ Efficient recovery of medium-high temperature heat by chemical recuperation of sensible heat ✓ Limited temperature variations in the FC thanks to the feeding of the FC with a low-CH_4 fuel, leading to a reduced SOFC internal reforming ✗ Higher air flow rate in the FC needed to keep a target FC gas exit temperature ✗ Higher capital cost than adiabatic pre-reforming
C		External reforming by auto thermal or fired tubular reforming	Excludes option D	✓ Avoids the need of steam reforming catalyst in the FC ✓ Minimal temperature variations in the FC thanks to the minimal CH_4 content in the FC fuel. ✗ Maximum air flow rate in the FC needed to keep a target FC gas exit temperature ✗ Maximum capital cost for fuel processing
D		Internal reforming	Excludes option C	✓ Efficient recovery of high temperature heat generated in the FC ✓ The higher the internal reforming, the lower the air flow rate to keep a given FC gas exit temperature thanks to the heat absorbed by reforming reaction. Lower air flow rates result in higher plant efficiency (higher combustion temperature of FC off-gases, lower stack losses) ✓ Simpler and more compact overall configuration because it avoids external heat exchangers and reactors ✗ Need of reforming catalyst on anode surface: higher FC cost ✗ Lower FC power density for given voltage, due to the kinetics of steam reforming limiting the fuel conversion process, especially in intermediate-low temperature FC. ✗ Higher temperature variations in the FC due to the endothermic nature of the steam reforming reaction
E	WGS	Pre-Anode WGS	Needs option C Excludes option F	✓ Risk of carbon deposition in the FC is minimized by feeding H_2 ✗ Increases plant complexity

F		Internal WGS	Excludes option E	✓ More compact and simpler configuration
G	Fuel humidification	Steam from boiler	Excludes options H, I	✓ Simple and easily controllable system ✗ Demi-water consumption, eventually emitted to the atmosphere in case of configuration with no water recovery
H		Anode recycle by blower	Excludes options G, I	✓ No additional water consumption ✗ Additional electric consumption
I		Anode recycle by ejector	Excludes options G, H	✓ No additional water consumption ✓ High temperature recycle possible, contributing to fuel preheating to high temperature ✓ No additional power consumption, if primary fuel is available at sufficiently high pressure and pressure energy is not recovered by a fuel expander. ✗ In case primary fuel is available at low pressure (or a fuel expander is used), this is a dissipative system, which requires a higher compression power (or leads to lower power generation from fuel expander) than a recycle blower.
J	Air final preheating	High temperature heat exchanger	Excludes option K	✓ Efficient system, recovering high temperature waste heat from the FC. ✗ Need of high temperature and high cost heat exchanger
K		Ejector driven cathode recycle	Excludes option J	✓ Cheap and compact system, not requiring high temperature heat exchange surfaces ✗ Dissipative system, lead to higher electric consumption for air compression
L		Atmospheric pressure FC with bottoming Rankine cycle	Excludes options M, N	✓ Maximum plant simplicity ✗ Lower cell performance because of the lower Nernst potential (for the same inlet composition and temperature) compared to alternative options.
M	Operating pressure and bottoming cycle	Moderate pressure (2-8 bar) FC with low T regenerative gas turbine cycle (uncooled turbine) and optional Rankine cycle	Excludes options L, N	✓ Increased FC performance thanks to the increased Nernst potential ✗ Increased plant complexity due to the need of balancing air and fuel channels' pressure and of containing vessel for the FC

N		High pressure (10-30 bar) FC with high b and high temperature gas turbine cycle (cooled turbine) and Rankine cycle	Excludes options L, M	✓ Maximum plant efficiency ✓ A relevant portion of the total power output is generated by the gas turbine cycle, which has a lower specific cost than the FC. This may have a positive effect on the overall plant Capex. ✗ Maximum plant complexity
O	CO_2 capture process	Oxycombustion of anode exhaust	Excludes options P, Q, R	✓ Relatively low O_2 consumption expected (reducing for increasing FC U_f factors) ✓ A virtually complete CO_2 capture efficiency is possible ✗ High energy consumption for oxygen production ✗ High cost of ASU, especially at small size, negatively affects overall plant CAPEX
P		WGS on anode exhaust and CO_2/H_2 separation	Excludes options O, Q, R Needs options S or T	✓ No need of further steam dilution, since water produced in the cell by hydrogen oxidation leads to high S/C ✓ H_2 can be recovered and efficiently converted by partial recycle to the FC or by combustion in the bottoming cycle (especially in case of a gas turbine cycle) ✗ CO_2 capture efficiency highly depends on the fuel utilisation in the SOFC
Q		Pre-Anode H_2 production and H_2/CO_2 separation	Needs options C, E, S Excludes options O, P, R	✓ Simple configuration in case pre-SOFC steam reforming and WGS are selected for optimal operation of the FC ✗ Solvent system needed for CO_2 absorption due to the relatively low CO_2 concentration ✗ Does not exploit the intrinsic capability of N_2 free fuel oxidation of SOFCs
R		Post-Anode chemical absorption on air combusted anode gas	Excludes options O, P, Q, S, T	✓ No effect on the base FC plant without CO_2 capture: easily retrofittable e Solvent system needed for CO_2 absorption due to the low CO_2 concentration ✗ Does not exploit the intrinsic capability of N_2 free fuel oxidation of FC ✗ Low energy efficiency expected due to the heat required for solvent regeneration
S	CO_2/H_2 separation process	Physical/chemical absorption	Needs options P or Q Excludes options O, R, T	✓ Do not need high CO_2 concentrations for efficient separation ✓ CO_2 is released at intrinsically high purity ✗ Heat required for chemical solvent regeneration ✗ Lower energy efficiency expected compared to low temperature phase separation, in case of high initial CO_2 concentrations

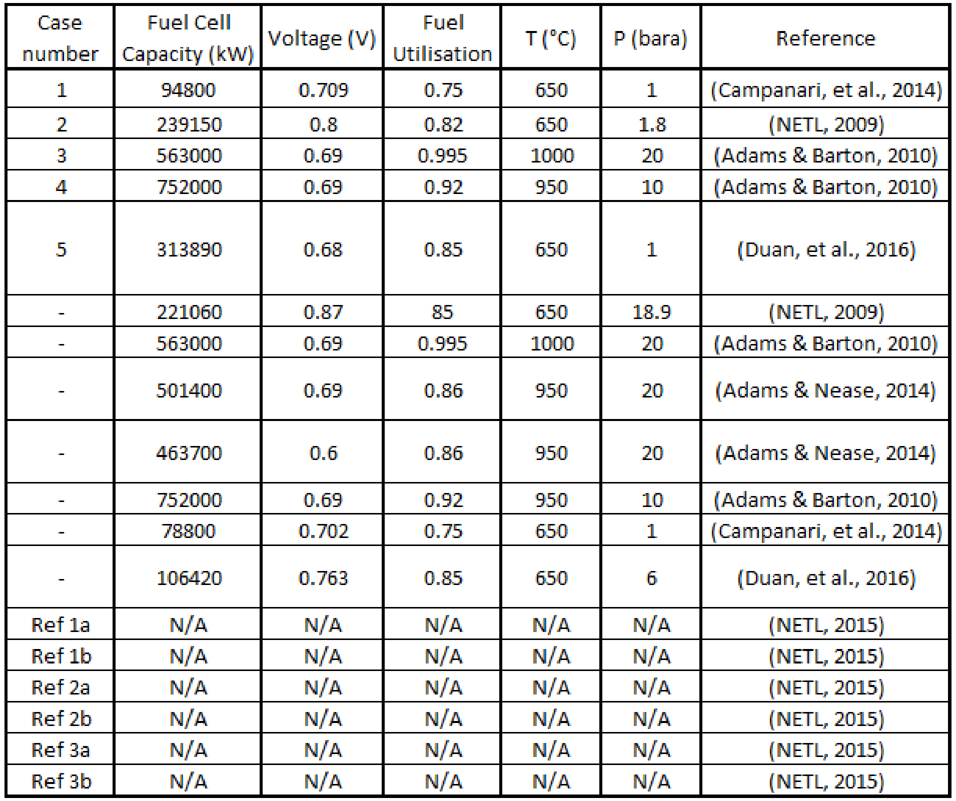
T		Low temperature phase separation	Needs option P Excludes option O, Q, R, S	✓ Efficient system in case of high initial CO_2 concentrations ✗ Needs high U_f factors

Appendix 3

Techno-economic Assessment Detailed Results

Case number	Location	Currency	Monetary Year	Fuel	General Technology	CO_2 Capture Technology	Gross Plant Output (kW)	Net Plant Output (kWe)	Reference
1	Netherlands	EUR	2008	Natural Gas	Atmospheric NGFC	Cryogenic CO_2 separation	530700	505900	(Campanari, et al., 2014)
2	USA (Mid-west)	USD	2007	Coal	Atmospheric IGFC	Oxycombustion followed by condensation	299470	253390	(NETL, 2009)
3	USA (Mid-west)	USD	2007	Coal	Pressurised IGFC	Oxidation unit followed by flash cascade	1061400	765500	(Adams & Barton, 2010)
4	USA (Mid-west)	USD	2007	Natural Gas	Pressurised NGFC	Oxidation unit followed by flash cascade	849800	693000	(Adams & Barton, 2010)
5	USA (Mid-west)	USD	2014	Anode: NG Cathode: Exhaust of coal fired pp	Coal Fired Power Plant + Atmospheric MCFC + Afterburner + HRSG ST	Oxycombustion followed by condensation	999290	935270	(Duan, et al., 2016)
-	USA (Mid-west)	USD	2007	Coal	Pressurised IGFC	Oxycombustion followed by condensation	347488	253000	(NETL, 2009)
-	USA (Mid-west)	USD	2007	Coal	Pressurised IGFC	Oxidation unit followed by flash cascade	1061400	765500	(Adams & Barton, 2010)
-	USA (Mid-west)	USD	2007	Coal	Pressurised IGFC	Oxidation unit followed by flash cascade	-	719000	(Adams & Nease, 2014)
-	USA (Mid-west)	USD	2007	Coal	Pressurised IGFC	Oxidation unit followed by flash cascade	-	719000	(Adams & Nease, 2014)
-	USA (Mid-west)	USD	2007	Natural Gas	Pressurised NGFC	Oxidation unit followed by flash cascade	849800	693000	(Adams & Barton, 2010)
-	Netherlands	EUR	2008	Natural Gas	Atmospheric NGFC	Oxy-combustion	533900	510900	(Campanari, et al., 2014)
-	USA (Mid-west)	USD	2014	Natural Gas	MCFC + Afterburner + HRSG ST	Oxyfuel + condensation	136220	129480	(Duan, et al., 2016)
Ref 1a	USA (Mid-west)	USD	2011	Coal	Supercritical PC	N/A	580000	550312	(NETL, 2015)
Ref 1b	USA (Mid-west)	USD	2011	Coal	Supercritical PC + CCS	Cansolv Amine Scrubbing	642000	550715	(NETL, 2015)
Ref 2a	USA (Mid-west)	USD	2011	Natural Gas	NGCC	N/A	641000	629928	(NETL, 2015)
Ref 2b	USA (Mid-west)	USD	2011	Natural Gas	NGCC + CCS	Cansolv Amine Scrubbing	601000	559268	(NETL, 2015)
Ref 3a	USA (Mid-west)	USD	2011	Coal	IGCC	N/A	748000	622140	(NETL, 2015)
Ref 3b	USA (Mid-west)	USD	2011	Coal	IGCC + CCS	Two-Stage Selexol	735000	544130	(NETL, 2015)

Table A3-1 – Summary of key parameters from all cases covered in techno-economic assessment (Part 1)



Case number	Fuel Cell Capacity (kW)	Voltage (V)	Fuel Utilisation	T (°C)	P (bara)	Reference
1	94800	0.709	0.75	650	1	(Campanari, et al., 2014)
2	239150	0.8	0.82	650	1.8	(NETL, 2009)
3	563000	0.69	0.995	1000	20	(Adams & Barton, 2010)
4	752000	0.69	0.92	950	10	(Adams & Barton, 2010)
5	313890	0.68	0.85	650	1	(Duan, et al., 2016)
-	221060	0.87	85	650	18.9	(NETL, 2009)
-	563000	0.69	0.995	1000	20	(Adams & Barton, 2010)
-	501400	0.69	0.86	950	20	(Adams & Nease, 2014)
-	463700	0.6	0.86	950	20	(Adams & Nease, 2014)
-	752000	0.69	0.92	950	10	(Adams & Barton, 2010)
-	78800	0.702	0.75	650	1	(Campanari, et al., 2014)
-	106420	0.763	0.85	650	6	(Duan, et al., 2016)
Ref 1a	N/A	N/A	N/A	N/A	N/A	(NETL, 2015)
Ref 1b	N/A	N/A	N/A	N/A	N/A	(NETL, 2015)
Ref 2a	N/A	N/A	N/A	N/A	N/A	(NETL, 2015)
Ref 2b	N/A	N/A	N/A	N/A	N/A	(NETL, 2015)
Ref 3a	N/A	N/A	N/A	N/A	N/A	(NETL, 2015)
Ref 3b	N/A	N/A	N/A	N/A	N/A	(NETL, 2015)

Table A3-1 – Summary of key parameters from all cases covered in techno-economic assessment (Part 2)

Case number	Currency format (Millions)	Coal and Sorbent Handling	Coal Prep and Feed System	Coal fired Power Plant	Catalyst Treatment	Ash Handling	Catalyst Recovery	Gasifier	Natural Gas Turbine	Natural Gas Expander	Natural Gas Preheating	Reformer Pre Heating	Pre-reformer	Reformer	ASU	Dry Gas / Humid Gas Cleaning
1	Euro	0.00	0.00	0.00	0.00	0.00	0.00	0.00	108.80	0.00	0.00	0.00	0.00	0.00	0.00	0.00
2	US $	11.00	59.00	0.00	6.00	14.00	10.00	37.00	0.00	0.00	0.00	0.00	0.00	0.00	38.00	31.00
3	US $	32.59	56.71	0.00	0.00	44.55	0.00	244.08	0.00	0.00	0.00	0.00	0.00	0.00	190.01	32.84
4	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.67	2.25	2.77	0.09	92.89	79.33	0.00
5	US $	0.00	0.00	477.45	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	25.97	0.00
-	US $	10.00	51.00	0.00	5.00	16.00	9.00	32.00	0.00	0.00	0.00	0.00	0.00	0.00	32.00	42.00
-	US $	32.59	56.71	0.00	0.00	44.55	0.00	244.08	0.00	0.00	0.00	0.00	0.00	0.00	190.01	32.84
-	US $	35.29	56.71	0.00	0.00	44.55	0.00	244.08	0.00	0.00	0.00	0.00	0.00	0.00	191.73	32.84
-	US $	37.25	59.86	0.00	0.00	47.02	0.00	257.62	0.00	0.00	0.00	0.00	0.00	0.00	202.37	34.66
-	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.67	2.25	2.77	0.09	92.89	79.33	0.00
-	Euro	0.00	0.00	0.00	0.00	0.00	0.00	0.00	108.80	0.00	0.00	0.00	0.00	0.00	13.66	0.00
-	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	6.17	0.00
Ref 1a	US $	45.40	21.53	341.72	0.00	16.78	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	167.27
Ref 1b	US $	52.29	24.98	400.79	0.00	19.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	197.48
Ref 2a	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	134.93	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 2b	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	134.93	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 3a	US $	43.25	69.31	0.00	0.00	54.15	0.00	590.40	144.44	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 3b	US $	44.40	71.29	0.00	0.00	55.56	0.00	646.65	160.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00

Table A3-2 – Summary of All Cases Installed Costs Pre-Rationalisation (Part 1)

Case number	Currency format (Millions)	Sulphur Polishing	Mecury Removal	Selexol Process	Misc. Cleaning	Syngas Expander	Fuel Cell	Fuel Cell Air Blower and Heat Exchanger	Fuel Cell Turbo Machinery	Air exhaust Expander	Exhaust Gas Blower	Oxy - Combustor
1	Euro	0.00	0.00	0.00	0.00	0.00	604.11	0.00	0.00	0.00	2.86	0.00
2	US $	13.00	0.00	0.00	0.00	2.00	146.00	20.00	0.00	0.00	0.00	5.00
3	US $	32.79	3.25	25.59	0.00	14.13	281.50	26.59	0.00	62.17	0.00	0.00
4	US $	0.00	0.00	0.00	0.00	0.00	751.69	17.27	0.00	25.72	0.00	0.00
5	US $	0.00	0.00	0.00	0.00	0.00	2499.92	0.00	0.00	0.00	0.00	0.00
-	US $	13.00	0.00	0.00	0.00	2.00	95.00	58.00	0.00	0.00	0.00	5.00
-	US $	32.79	3.25	25.59	0.00	14.13	563.00	26.59	0.00	62.17	0.00	0.00
-	US $	32.79	3.25	25.59	6.68	10.75	501.40	103.74	109.09	36.39	0.00	0.00
-	US $	34.61	3.43	27.01	7.05	11.34	463.70	109.49	115.14	38.41	0.00	0.00
-	US $	0.00	0.00	0.00	0.00	0.00	375.84	17.27	0.00	25.72	0.00	0.00
-	Euro	0.00	0.00	0.00	0.00	0.00	502.17	0.00	0.00	0.00	2.86	0.00
-	US $	0.00	0.00	0.00	0.00	0.00	287.82	3.81	7.81	0.00	0.00	0.00
Ref 1a	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 1b	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 2a	US $	0.00	0.00	0.00	4.96	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 2b	US $	0.00	0.00	0.00	4.96	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 3a	US $	39.62	3.67	123.04	16.90	10.28	0.00	0.00	0.00	0.00	0.00	0.00
Ref 3b	US $	40.70	4.78	251.15	24.90	9.63	0.00	0.00	0.00	0.00	0.00	0.00

Table A3-2 – Summary of All Cases Installed Costs Pre-Rationalisation (Part 2)

Case number	Currency format (Millions)	HRSG	Steam Turbine	CO_2 Capture	CO_2 & H_2O Separation	Water Gas Shift Unit	CO_2 Compression	Cooling Water System	Boiler Plant and Water	Feed water and Misc. BoP	Condenser and Accesserories	Stack	Feed water steam generator	Initial Charge of solvents, catalyst etc.
1	Euro	66.51	64.31	0.00	0.00	0.00	24.31	78.85	0.00	17.18	0.00	0.00	0.00	0.00
2	US $	13.00	14.00	0.00	0.00	0.00	11.00	8.00	0.00	8.00	0.00	0.00	0.00	0.00
3	US $	0.00	60.95	0.00	5.96	15.20	38.77	34.90	11.71	33.17	0.00	0.00	59.27	3.10
4	US $	0.00	29.89	0.00	4.74	8.10	26.40	33.01	0.00	46.36	3.22	3.23	36.45	0.00
5	US $	46.56	0.00	0.00	0.00	0.00	70.70	0.00	0.00	0.00	0.00	0.00	0.00	0.00
-	US $	0.00	0.00	0.00	0.00	0.00	11.00	10.00	0.00	0.00	0.00	0.00	0.00	0.00
-	US $	0.00	60.95	0.00	5.96	15.20	38.77	34.90	11.71	33.17	0.00	0.00	59.27	3.10
-	US $	0.00	63.40	0.00	2.30	13.67	47.93	38.15	16.37	37.04	10.72	3.23	53.85	2.99
-	US $	0.00	68.59	0.00	2.43	0.00	50.59	40.26	17.28	39.09	11.60	3.41	58.26	3.15
-	US $	0.00	29.89	0.00	4.74	8.10	26.40	33.01	0.00	46.36	3.22	3.23	36.45	0.00
-	Euro	73.34	69.82	0.00	0.00	0.00	22.89	82.59	0.00	28.85	0.00	0.00	0.00	0.00
-	US $	6.79	6.79	0.00	0.00	0.00	5.79	0.00	0.00	15.09	0.00	0.00	0.00	0.00
Ref 1a	US $	0.00	166.93	0.00	0.00	0.00	0.00	44.04	0.00	93.64	0.00	45.63	0.00	0.00
Ref 1b	US $	0.00	178.18	632.14	0.00	0.00	0.00	62.25	0.00	112.15	0.00	45.03	0.00	0.00
Ref 2a	US $	46.96	81.12	0.00	0.00	0.00	0.00	19.91	0.00	53.87	0.00	1.40	0.00	0.00
Ref 2b	US $	44.24	74.54	378.18	0.00	0.00	0.00	27.50	0.00	57.94	0.00	1.12	0.00	0.00
Ref 3a	US $	44.89	96.40	0.00	0.00	0.00	0.00	37.89	0.00	66.67	0.00	13.53	0.00	0.00
Ref 3b	US $	42.69	93.12	0.00	0.00	0.00	84.09	40.59	0.00	69.04	0.00	13.72	0.00	0.00

Table A3-2 – Summary of All Cases Installed Costs Pre-Rationalisation (Part 3)

Case number	Currency format (Millions)	Piping	Accessory Electric Plant	Instrumentation and Control	Site Improvements	Building and Structures	Equipment Installation	Engineering and Supervision	Construction Expenses	Legal Expenses	Contractors Fee	Contingency	Total Plant Cost
1	Euro	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	966.94
2	US $	0.00	31.00	10.00	8.00	8.00	0.00	0.00	0.00	0.00	0.00	0.00	503.00
3	US $	0.00	89.78	25.94	19.31	17.58	0.00	0.00	0.00	0.00	0.00	0.00	1462.41
4	US $	0.00	36.23	17.23	9.06	9.89	0.00	0.00	0.00	0.00	0.00	0.00	1239.48
5	US $	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3120.61
-	US $	0.00	27.00	9.00	7.00	7.00	0.00	0.00	0.00	0.00	0.00	0.00	441.00
-	US $	0.00	89.78	25.94	19.31	17.58	0.00	0.00	0.00	0.00	0.00	0.00	1743.91
-	US $	0.00	85.06	25.94	19.31	17.58	0.00	0.00	0.00	0.00	0.00	0.00	1872.40
-	US $	0.00	89.78	27.38	20.38	18.56	0.00	0.00	0.00	0.00	0.00	0.00	1899.73
-	US $	0.00	36.23	17.23	9.06	9.89	0.00	0.00	0.00	0.00	0.00	0.00	863.64
-	Euro	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	904.98
-	US $	231.25	37.41	122.43	34.01	95.22	159.83	112.22	139.43	13.60	74.82	149.63	1509.91
Ref 1a	US $	0.00	61.74	26.32	16.39	66.97	0.00	0.00	0.00	0.00	0.00	0.00	1114.36
Ref 1b	US $	0.00	93.58	31.65	18.06	71.53	0.00	0.00	0.00	0.00	0.00	0.00	1939.14
Ref 2a	US $	0.00	45.62	16.86	11.77	13.53	0.00	0.00	0.00	0.00	0.00	0.00	430.93
Ref 2b	US $	0.00	59.81	19.57	11.99	13.13	0.00	0.00	0.00	0.00	0.00	0.00	827.90
Ref 3a	US $	0.00	93.73	30.04	22.71	22.12	0.00	0.00	0.00	0.00	0.00	0.00	1523.05
Ref 3b	US $	0.00	107.85	33.43	23.64	22.57	0.00	0.00	0.00	0.00	0.00	0.00	1840.12

Table A3-2 – Summary of All Cases Installed Costs Pre-Rationalisation (Part 4)

Basis - Netherlands, 2017, Million Euros

Case number	Contingency Factor	Material Factor	Labour Factor (Productivity)	Labour Factor Cost	Equipment Factor	Currency Conversion	Currency Inflation Factor	Coal and Sorbent Handling	Coal Prep and Feed System	Coal fired Power Plant	Catalyst Treatment	Ash Handling
1	1.000	1.000	1.000	1.000	0.921	1.000	1.126	0.0	0.0	0.0	0.0	0.0
2	1.000	1.064	0.950	1.087	1.076	1.370	1.154	17.1	91.6	0.0	9.3	21.7
3	1.000	1.064	0.950	1.087	1.076	1.370	1.154	50.6	88.0	0.0	0.0	69.2
4	1.000	1.064	0.950	1.087	1.076	1.370	1.154	0.0	0.0	0.0	0.0	0.0
5	1.000	1.064	0.950	1.087	0.955	1.329	1.012	0.0	0.0	634.2	0.0	0.0
-	1.000	1.064	0.950	1.087	1.076	1.370	1.154	15.5	79.2	0.0	7.8	24.8
-	1.000	1.064	0.950	1.087	1.076	1.370	1.154	50.6	88.0	0.0	0.0	69.2
-	1.000	1.064	0.950	1.087	1.076	1.370	1.154	54.8	88.0	0.0	0.0	69.2
-	1.000	1.064	0.950	1.087	1.076	1.370	1.154	57.8	92.9	0.0	0.0	73.0
-	1.000	1.064	0.950	1.087	1.076	1.370	1.154	0.0	0.0	0.0	0.0	0.0
-	1.000	1.000	1.000	1.000	0.921	1.000	1.126	0.0	0.0	0.0	0.0	0.0
-	1.000	1.064	0.950	1.087	0.955	1.329	1.012	0.0	0.0	0.0	0.0	0.0
Ref 1a	1.000	1.064	0.950	1.087	0.928	1.393	1.074	64.3	30.5	484.1	0.0	23.8
Ref 1b	1.000	1.064	0.950	1.087	0.928	1.393	1.074	74.1	35.4	567.8	0.0	27.0
Ref 2a	1.000	1.064	0.950	1.087	0.928	1.393	1.074	0.0	0.0	0.0	0.0	0.0
Ref 2b	1.000	1.064	0.950	1.087	0.928	1.393	1.074	0.0	0.0	0.0	0.0	0.0
Ref 3a	1.000	1.064	0.950	1.087	0.928	1.393	1.074	61.3	98.2	0.0	0.0	76.7
Ref 3b	1.000	1.064	0.950	1.087	0.928	1.393	1.074	62.9	101.0	0.0	0.0	78.7

Table A3-3 – Summary of All Cases Installed Costs Post-Rationalisation (Part 1)

Basis - Netherlands, 2017, Million Euros

Case number	Catalyst Recovery	Gasifier	Natural Gas Turbine	Natural Gas Expander	Natural Gas Preheating	Reformer Pre Heating	Pre-reformer	Reformer	ASU	Dry Gas / Humid Gas Cleaning	Sulphur Polishing	Mecury Removal	Selexol Process	Misc. Cleaning	Syngas Expander
1	0.0	0.0	111.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
2	15.5	57.4	0.0	0.0	0.0	0.0	0.0	0.0	59.0	48.1	20.2	0.0	0.0	0.0	3.1
3	0.0	378.9	0.0	0.0	0.0	0.0	0.0	0.0	295.0	51.0	50.9	5.0	39.7	0.0	21.9
4	0.0	0.0	0.0	5.7	3.5	4.3	0.1	144.2	123.2	0.0	0.0	0.0	0.0	0.0	0.0
5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	34.5	0.0	0.0	0.0	0.0	0.0	0.0
-	14.0	49.7	0.0	0.0	0.0	0.0	0.0	0.0	49.7	65.2	20.2	0.0	0.0	0.0	3.1
-	0.0	378.9	0.0	0.0	0.0	0.0	0.0	0.0	295.0	51.0	50.9	5.0	39.7	0.0	21.9
-	0.0	378.9	0.0	0.0	0.0	0.0	0.0	0.0	297.7	51.0	50.9	5.0	39.7	10.4	16.7
-	0.0	400.0	0.0	0.0	0.0	0.0	0.0	0.0	314.2	53.8	53.7	5.3	41.9	11.0	17.6
-	0.0	0.0	0.0	5.7	3.5	4.3	0.1	144.2	123.2	0.0	0.0	0.0	0.0	0.0	0.0
-	0.0	0.0	111.3	0.0	0.0	0.0	0.0	0.0	14.0	0.0	0.0	0.0	0.0	0.0	0.0
-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	8.2	0.0	0.0	0.0	0.0	0.0	0.0
Ref 1a	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	237.0	0.0	0.0	0.0	0.0	0.0
Ref 1b	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	279.8	0.0	0.0	0.0	0.0	0.0
Ref 2a	0.0	0.0	191.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	7.0	0.0
Ref 2b	0.0	0.0	191.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	7.0	0.0
Ref 3a	0.0	836.4	204.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	56.1	5.2	174.3	23.9	14.6
Ref 3b	0.0	916.1	227.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	57.7	6.8	355.8	35.3	13.6

Table A3-3 – Summary of All Cases Installed Costs Post-Rationalisation (Part 2)

Basis - Netherlands, 2017, Million Euros

Case number	Fuel Cell	Fuel Cell Air Blower and Heat Exchanger	Fuel Cell Turbo Machinery	Air exhaust Expander	Exhaust Gas Blower	Oxy - Combustor	HRSG	Steam Turbine	CO$_2$ Capture	CO$_2$ & H$_2$O Separation	Water Gas Shift Unit
1	618.0	0.0	0.0	0.0	2.9	0.0	68.0	65.8	0.0	0.0	0.0
2	226.7	31.1	0.0	0.0	0.0	7.8	20.2	21.7	0.0	0.0	0.0
3	437.0	41.3	0.0	96.5	0.0	0.0	0.0	94.6	0.0	9.3	23.6
4	1167.0	26.8	0.0	39.9	0.0	0.0	0.0	46.4	0.0	7.4	12.6
5	3320.5	0.0	0.0	0.0	0.0	0.0	61.8	0.0	0.0	0.0	0.0
-	147.5	90.0	0.0	0.0	0.0	7.8	0.0	0.0	0.0	0.0	0.0
-	874.1	41.3	0.0	96.5	0.0	0.0	0.0	94.6	0.0	9.3	23.6
-	778.5	161.1	169.4	56.5	0.0	0.0	0.0	98.4	0.0	3.6	21.2
-	719.9	170.0	178.8	59.6	0.0	0.0	0.0	106.5	0.0	3.8	0.0
-	583.5	26.8	0.0	39.9	0.0	0.0	0.0	46.4	0.0	7.4	12.6
-	513.7	0.0	0.0	0.0	2.9	0.0	75.0	71.4	0.0	0.0	0.0
-	382.3	5.1	10.4	0.0	0.0	0.0	9.0	9.0	0.0	0.0	0.0
Ref 1a	0.0	0.0	0.0	0.0	0.0	0.0	0.0	236.5	0.0	0.0	0.0
Ref 1b	0.0	0.0	0.0	0.0	0.0	0.0	0.0	252.4	895.5	0.0	0.0
Ref 2a	0.0	0.0	0.0	0.0	0.0	0.0	66.5	114.9	0.0	0.0	0.0
Ref 2b	0.0	0.0	0.0	0.0	0.0	0.0	62.7	105.6	535.8	0.0	0.0
Ref 3a	0.0	0.0	0.0	0.0	0.0	0.0	63.6	136.6	0.0	0.0	0.0
Ref 3b	0.0	0.0	0.0	0.0	0.0	0.0	60.5	131.9	0.0	0.0	0.0

Table A3-3 – Summary of All Cases Installed Costs Post-Rationalisation (Part 3)

Basis - Netherlands, 2017, Million Euros

Case number	Water Gas Shift Unit	CO_2 Compression	Cooling Water System	Boiler Plant and Water	Feed water and Misc. BoP	Condenser and Accesserories	Stack	Feed water steam generator	Initial Charge of solvents, catalyst etc.	Piping	Accessory Electric Plant
1	0.0	24.9	80.7	0.0	17.6	0.0	0.0	0.0	0.0	0.0	0.0
2	0.0	17.1	12.4	0.0	12.4	0.0	0.0	0.0	0.0	0.0	48.1
3	23.6	60.2	54.2	18.2	51.5	0.0	0.0	92.0	4.8	0.0	139.4
4	12.6	41.0	51.3	0.0	72.0	5.0	5.0	56.6	0.0	0.0	56.2
5	0.0	93.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
-	0.0	17.1	15.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	41.9
-	23.6	60.2	54.2	18.2	51.5	0.0	0.0	92.0	4.8	0.0	139.4
-	21.2	74.4	59.2	25.4	57.5	16.6	5.0	83.6	4.6	0.0	132.1
-	0.0	78.5	62.5	26.8	60.7	18.0	5.3	90.5	4.9	0.0	139.4
-	12.6	41.0	51.3	0.0	72.0	5.0	5.0	56.6	0.0	0.0	56.2
-	0.0	23.4	84.5	0.0	29.5	0.0	0.0	0.0	0.0	0.0	0.0
-	0.0	7.7	0.0	0.0	20.0	0.0	0.0	0.0	0.0	307.2	49.7
Ref 1a	0.0	0.0	62.4	0.0	132.7	0.0	64.6	0.0	0.0	0.0	87.5
Ref 1b	0.0	0.0	88.2	0.0	158.9	0.0	63.8	0.0	0.0	0.0	132.6
Ref 2a	0.0	0.0	28.2	0.0	76.3	0.0	2.0	0.0	0.0	0.0	64.6
Ref 2b	0.0	0.0	39.0	0.0	82.1	0.0	1.6	0.0	0.0	0.0	84.7
Ref 3a	0.0	0.0	53.7	0.0	94.5	0.0	19.2	0.0	0.0	0.0	132.8
Ref 3b	0.0	119.1	57.5	0.0	97.8	0.0	19.4	0.0	0.0	0.0	152.8

Table A3-3 – Summary of All Cases Installed Costs Post-Rationalisation (Part 4)

Basis - Netherlands, 2017, Million Euros

Case number	Instrumentation and Control	Site Improvements	Building and Structures	Equipment Installation	Engineering and Supervision	Construction Expenses	Legal Expenses	Contractors Fee	Contingency	Total Plant Cost
1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	989.2
2	15.5	12.4	12.4	0.0	0.0	0.0	0.0	0.0	0.0	780.9
3	40.3	30.0	27.3	0.0	0.0	0.0	0.0	0.0	0.0	2270.5
4	26.7	14.1	15.3	0.0	0.0	0.0	0.0	0.0	0.0	1924.4
5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	4144.9
-	14.0	10.9	10.9	0.0	0.0	0.0	0.0	0.0	0.0	684.7
-	40.3	30.0	27.3	0.0	0.0	0.0	0.0	0.0	0.0	2707.5
-	40.3	30.0	27.3	0.0	0.0	0.0	0.0	0.0	0.0	2907.0
-	42.5	31.6	28.8	0.0	0.0	0.0	0.0	0.0	0.0	2949.5
-	26.7	14.1	15.3	0.0	0.0	0.0	0.0	0.0	0.0	1340.9
-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	925.8
-	162.6	45.2	126.5	212.3	149.1	185.2	18.1	99.4	198.7	2005.5
Ref 1a	37.3	23.2	94.9	0.0	0.0	0.0	0.0	0.0	0.0	1578.7
Ref 1b	44.8	25.6	101.3	0.0	0.0	0.0	0.0	0.0	0.0	2747.2
Ref 2a	23.9	16.7	19.2	0.0	0.0	0.0	0.0	0.0	0.0	610.5
Ref 2b	27.7	17.0	18.6	0.0	0.0	0.0	0.0	0.0	0.0	1172.9
Ref 3a	42.6	32.2	31.3	0.0	0.0	0.0	0.0	0.0	0.0	2157.7
Ref 3b	47.4	33.5	32.0	0.0	0.0	0.0	0.0	0.0	0.0	2606.9

Table A3-3 – Summary of All Cases Installed Costs Post-Rationalisation (Part 5)

Basis - Netherlands, 2017, Euros per kW

TEA Case number	Coal and Sorbent Handling	Coal Prep and Feed System	Coal fired Power Plant	Catalyst Treatment	Ash Handling	Catalyst Recovery	Gasifier	Natural Gas Turbine	Natural Gas Expander	Natural Gas Preheating	Reformer Pre Heating	Pre-reformer	Reformer	ASU	Dry Gas / Humid Gas Cleaning
1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	220.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
2	67.4	361.5	0.0	36.8	85.8	61.3	226.7	0.0	0.0	0.0	0.0	0.0	0.0	232.8	189.9
3	66.1	115.0	0.0	0.0	90.3		494.9	0.0	0.0	0.0	0.0	0.0	0.0	385.3	66.6
4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	8.2	5.0	6.2	0.2	208.2	177.8	0.0
5	0.0	0.0	678.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	36.9	0.0
-	61.4	313.0	0.0	30.7	98.2	55.2	196.4	0.0	0.0	0.0	0.0	0.0	0.0	196.4	257.7
-	66.1	115.0	0.0	0.0	90.3	0.0	494.9	0.0	0.0	0.0	0.0	0.0	0.0	385.3	66.6
-	76.2	122.5	0.0	0.0	96.2	0.0	527.0	0.0	0.0	0.0	0.0	0.0	0.0	414.0	70.9
-	80.4	129.3	0.0	0.0	101.5	0.0	556.3	0.0	0.0	0.0	0.0	0.0	0.0	437.0	74.8
-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	8.2	5.0	6.2	0.2	208.2	177.8	0.0
-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	217.9	0.0	0.0	0.0	0.0	0.0	27.3	0.0
-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	63.3	0.0
Ref 1a	116.9	55.4	879.7	0.0	43.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	430.6
Ref 1b	134.5	64.3	1031.0	0.0	48.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	508.0
Ref 2a	0.0	0.0	0.0	0.0	0.0	0.0	0.0	303.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 2b	0.0	0.0	0.0	0.0	0.0	0.0	0.0	341.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 3a	97.3	155.9	0.0	0.0	121.8	0.0	1327.8	324.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 3b	115.6	185.6	0.0	0.0	144.7	0.0	1683.6	417.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Table A3-4 – Summary of All Cases Installed Costs per kW Post-Rationalisation (Part 1)

Basis - Netherlands, 2017, Euros per kW

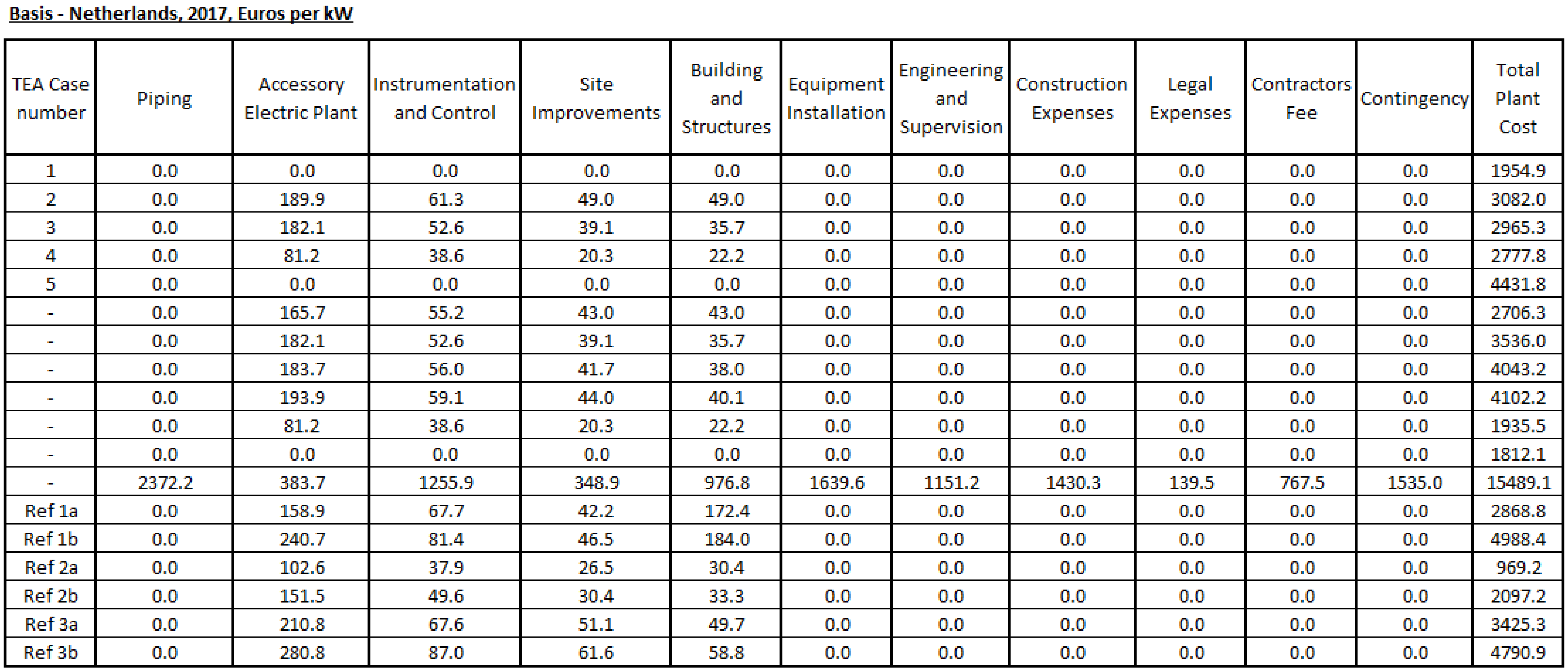
TEA Case number	Sulphur Polishing	Mecury Removal	Selexol Process	Misc. Cleaning	Syngas Expander	Fuel Cell	Fuel Cell Air Blower and Heat Exchanger	Fuel Cell Turbo Machinery	Air exhaust Expander	Exhaust Gas Blower	Oxy - Combustor
1	0.0	0.0	0.0	0.0	0.0	1221.3	0.0	0.0	0.0	5.8	0.0
2	79.7			0.0	12.3	894.6	122.5	0.0	0.0	0.0	30.6
3	66.5	6.6	51.9	0.0	28.7	570.8	53.9	0.0	126.0	0.0	0.0
4	0.0	0.0	0.0	0.0	0.0	1684.6	38.7	0.0	57.6	0.0	0.0
5	0.0	0.0	0.0	0.0	0.0	3550.3	0.0	0.0	0.0	0.0	0.0
-	79.8	0.0	0.0	0.0	12.3	583.0	355.9	0.0	0.0	0.0	30.7
-	66.5	6.6	51.9	0.0	28.7	1141.6	53.9	0.0	126.0	0.0	0.0
-	70.8	7.0	55.3	14.4	23.2	1082.7	224.0	235.6	78.6	0.0	0.0
-	74.7	7.4	58.3	15.2	24.5	1001.3	236.4	248.6	82.9	0.0	0.0
-	0.0	0.0	0.0	0.0	0.0	842.3	38.7	0.0	57.6	0.0	0.0
-	0.0	0.0	0.0	0.0	0.0	1005.5	0.0	0.0	0.0	5.7	0.0
-	0.0	0.0	0.0	0.0	0.0	2952.5	39.1	80.1	0.0	0.0	0.0
Ref 1a	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 1b	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 2a	0.0	0.0	0.0	11.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 2b	0.0	0.0	0.0	12.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ref 3a	89.1	8.3	276.7	38.0	23.1	0.0	0.0	0.0	0.0	0.0	0.0
Ref 3b	106.0	12.4	653.9	64.8	25.1	0.0	0.0	0.0	0.0	0.0	0.0

Table A3-4 – Summary of All Cases Installed Costs per kW Post-Rationalisation (Part 2)

Basis - Netherlands, 2017, Euros per kW

TEA Case number	HRSG	Steam Turbine	CO_2 Capture	CO_2 & H_2O Separation	Water Gas Shift Unit	CO_2 Compression	Cooling Water System	Boiler Plant and Water	Feed water and Misc. BoP	Condenser and Accesserories	Stack	Feed water steam generator	Initial Charge of solvents, catalyst etc.
1	134.5	130.0	0.0	0.0	0.0	49.1	159.4	0.0	34.7	0.0	0.0	0.0	0.0
2	79.7	85.8	0.0	0.0	0.0	67.4	49.0	0.0	49.0	0.0	0.0	0.0	0.0
3	0.0	123.6	0.0	12.1	30.8	78.6	70.8	23.7	67.2	0.0	0.0	120.2	6.3
4	0.0	67.0	0.0	10.6	18.2	59.2	74.0	0.0	103.9	7.2	7.2	81.7	0.0
5	66.1	0.0	0.0	0.0	0.0	100.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0
-	0.0	0.0	0.0	0.0	0.0	67.5	61.4	0.0	0.0	0.0	0.0	0.0	0.0
-	0.0	123.6	0.0	12.1	30.8	78.6	70.8	23.7	67.2	0.0	0.0	120.2	6.3
-	0.0	136.9	0.0	5.0	29.5	103.5	82.4	35.4	80.0	23.2	7.0	116.3	6.5
-	0.0	148.1	0.0	5.2		109.2	86.9	37.3	84.4	25.1	7.4	125.8	6.8
-	0.0	67.0	0.0	10.6	18.2	59.2	74.0	0.0	103.9	7.2	7.2	81.7	0.0
-	146.9	139.8	0.0	0.0	0.0	45.8	165.4	0.0	57.8	0.0	0.0	0.0	0.0
-	69.7	69.7	0.0	0.0	0.0	59.4	0.0	0.0	154.8	0.0	0.0	0.0	0.0
Ref 1a	0.0	429.7	0.0	0.0	0.0	0.0	113.4	0.0	241.1	0.0	117.5	0.0	0.0
Ref 1b	0.0	458.4	1626.2	0.0	0.0	0.0	160.1	0.0	288.5	0.0	115.8	0.0	0.0
Ref 2a	105.6	182.4	0.0	0.0	0.0	0.0	44.8	0.0	121.2	0.0	3.1	0.0	0.0
Ref 2b	112.1	188.8	958.0	0.0	0.0	0.0	69.7	0.0	146.8	0.0	2.8	0.0	0.0
Ref 3a	101.0	216.8	0.0	0.0	0.0	0.0	85.2	0.0	149.9	0.0	30.4	0.0	0.0
Ref 3b	111.2	242.4	0.0	0.0	0.0	218.9	105.7	0.0	179.7	0.0	35.7	0.0	0.0

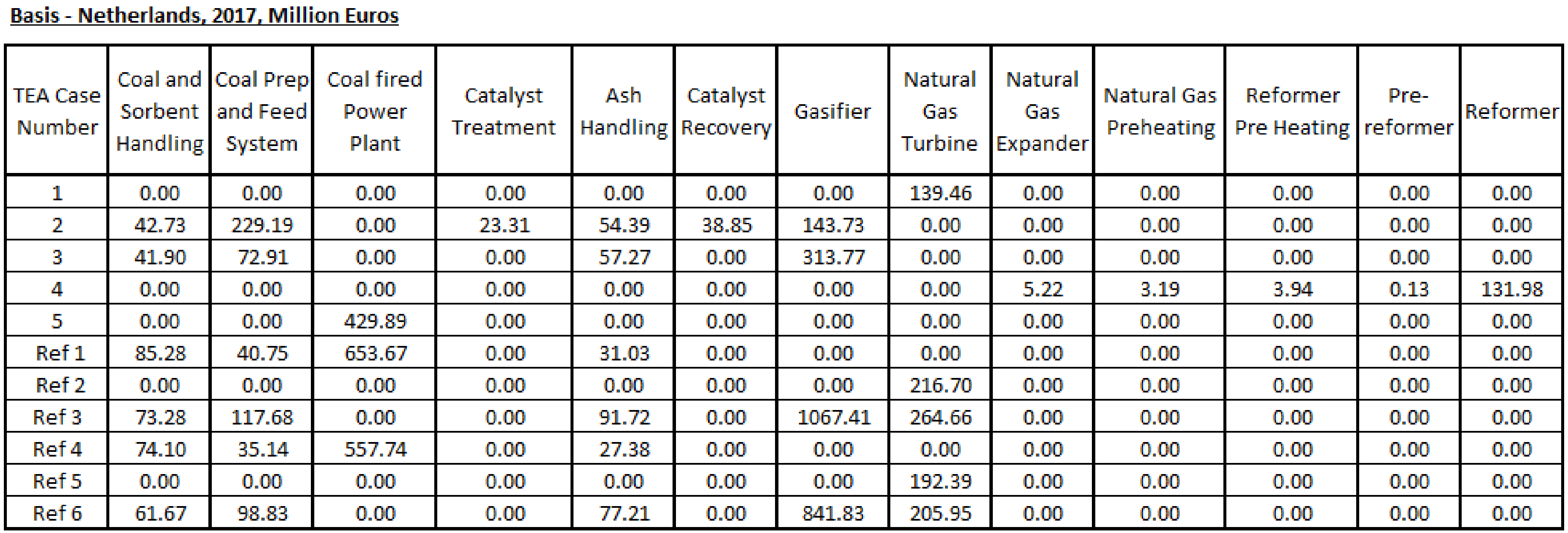
Table A3-4 – Summary of All Cases Installed Costs per kW Post-Rationalisation (Part 3)



Basis - Netherlands, 2017, Euros per kW

TEA Case number	Piping	Accessory Electric Plant	Instrumentation and Control	Site Improvements	Building and Structures	Equipment Installation	Engineering and Supervision	Construction Expenses	Legal Expenses	Contractors Fee	Contingency	Total Plant Cost
1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1954.9
2	0.0	189.9	61.3	49.0	49.0	0.0	0.0	0.0	0.0	0.0	0.0	3082.0
3	0.0	182.1	52.6	39.1	35.7	0.0	0.0	0.0	0.0	0.0	0.0	2965.3
4	0.0	81.2	38.6	20.3	22.2	0.0	0.0	0.0	0.0	0.0	0.0	2777.8
5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	4431.8
-	0.0	165.7	55.2	43.0	43.0	0.0	0.0	0.0	0.0	0.0	0.0	2706.3
-	0.0	182.1	52.6	39.1	35.7	0.0	0.0	0.0	0.0	0.0	0.0	3536.0
-	0.0	183.7	56.0	41.7	38.0	0.0	0.0	0.0	0.0	0.0	0.0	4043.2
-	0.0	193.9	59.1	44.0	40.1	0.0	0.0	0.0	0.0	0.0	0.0	4102.2
-	0.0	81.2	38.6	20.3	22.2	0.0	0.0	0.0	0.0	0.0	0.0	1935.5
-	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1812.1
-	2372.2	383.7	1255.9	348.9	976.8	1639.6	1151.2	1430.3	139.5	767.5	1535.0	15489.1
Ref 1a	0.0	158.9	67.7	42.2	172.4	0.0	0.0	0.0	0.0	0.0	0.0	2868.8
Ref 1b	0.0	240.7	81.4	46.5	184.0	0.0	0.0	0.0	0.0	0.0	0.0	4988.4
Ref 2a	0.0	102.6	37.9	26.5	30.4	0.0	0.0	0.0	0.0	0.0	0.0	969.2
Ref 2b	0.0	151.5	49.6	30.4	33.3	0.0	0.0	0.0	0.0	0.0	0.0	2097.2
Ref 3a	0.0	210.8	67.6	51.1	49.7	0.0	0.0	0.0	0.0	0.0	0.0	3425.3
Ref 3b	0.0	280.8	87.0	61.6	58.8	0.0	0.0	0.0	0.0	0.0	0.0	4790.9

Table A3-4 – Summary of All Cases Installed Costs per kW Post-Rationalisation (Part 4)



Basis - Netherlands, 2017, Million Euros

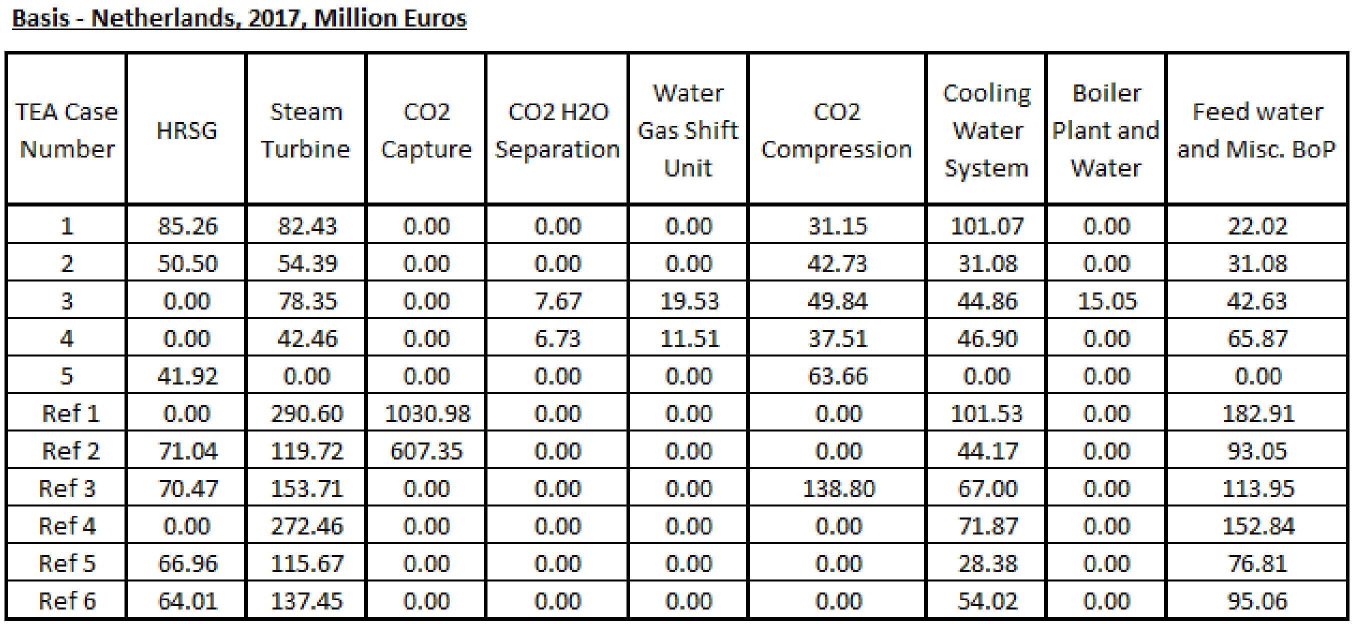
TEA Case Number	Coal and Sorbent Handling	Coal Prep and Feed System	Coal fired Power Plant	Catalyst Treatment	Ash Handling	Catalyst Recovery	Gasifier	Natural Gas Turbine	Natural Gas Expander	Natural Gas Preheating	Reformer Pre Heating	Pre-reformer	Reformer
1	0.00	0.00	0.00	0.00	0.00	0.00	0.00	139.46	0.00	0.00	0.00	0.00	0.00
2	42.73	229.19	0.00	23.31	54.39	38.85	143.73	0.00	0.00	0.00	0.00	0.00	0.00
3	41.90	72.91	0.00	0.00	57.27	0.00	313.77	0.00	0.00	0.00	0.00	0.00	0.00
4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	5.22	3.19	3.94	0.13	131.98
5	0.00	0.00	429.89	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 1	85.28	40.75	653.67	0.00	31.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	216.70	0.00	0.00	0.00	0.00	0.00
Ref 3	73.28	117.68	0.00	0.00	91.72	0.00	1067.41	264.66	0.00	0.00	0.00	0.00	0.00
Ref 4	74.10	35.14	557.74	0.00	27.38	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 5	0.00	0.00	0.00	0.00	0.00	0.00	0.00	192.39	0.00	0.00	0.00	0.00	0.00
Ref 6	61.67	98.83	0.00	0.00	77.21	0.00	841.83	205.95	0.00	0.00	0.00	0.00	0.00

Table A3-5 – Summary of All Cases Installed Costs Post-Rationalisation and Scaling (Part 1)

Basis - Netherlands, 2017, Million Euros

TEA Case Number	ASU	Dry Gas / Humid Gas Cleaning	Sulphur Polishing	Mecury Removal	One Stage Selexol	Misc. Cleaning	Syngas Expander	Fuel Cell	Fuel Cell Air Blower and Heat Exchanger	Air exhaust Expander	Exhaust Gas Blower	Oxy - Combustor
1	0.00	0.00	0.00	0.00	0.00	0.00	0.00	415.73	0.00	0.00	3.67	0.00
2	147.62	120.42	50.50	0.00	0.00	0.00	7.77	2094.30	77.69	0.00	0.00	19.42
3	244.26	42.21	42.15	4.17	32.89	0.00	18.16	2039.47	34.18	79.92	0.00	0.00
4	112.72	0.00	0.00	0.00	0.00	0.00	0.00	3010.94	24.54	36.55	0.00	0.00
5	23.38	0.00	0.00	0.00	0.00	0.00	0.00	744.73	0.00	0.00	0.00	0.00
Ref 1	0.00	322.07	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 2	0.00	0.00	0.00	0.00	0.00	7.97	0.00	0.00	0.00	0.00	0.00	0.00
Ref 3	0.00	0.00	67.18	7.89	414.56	41.11	15.89	0.00	0.00	0.00	0.00	0.00
Ref 4	0.00	273.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ref 5	0.00	0.00	0.00	0.00	0.00	7.07	0.00	0.00	0.00	0.00	0.00	0.00
Ref 6	0.00	0.00	56.49	5.24	175.44	24.10	14.66	0.00	0.00	0.00	0.00	0.00

Table A3-5 – Summary of All Cases Installed Costs Post-Rationalisation and Scaling (Part 2)



Basis - Netherlands, 2017, Million Euros

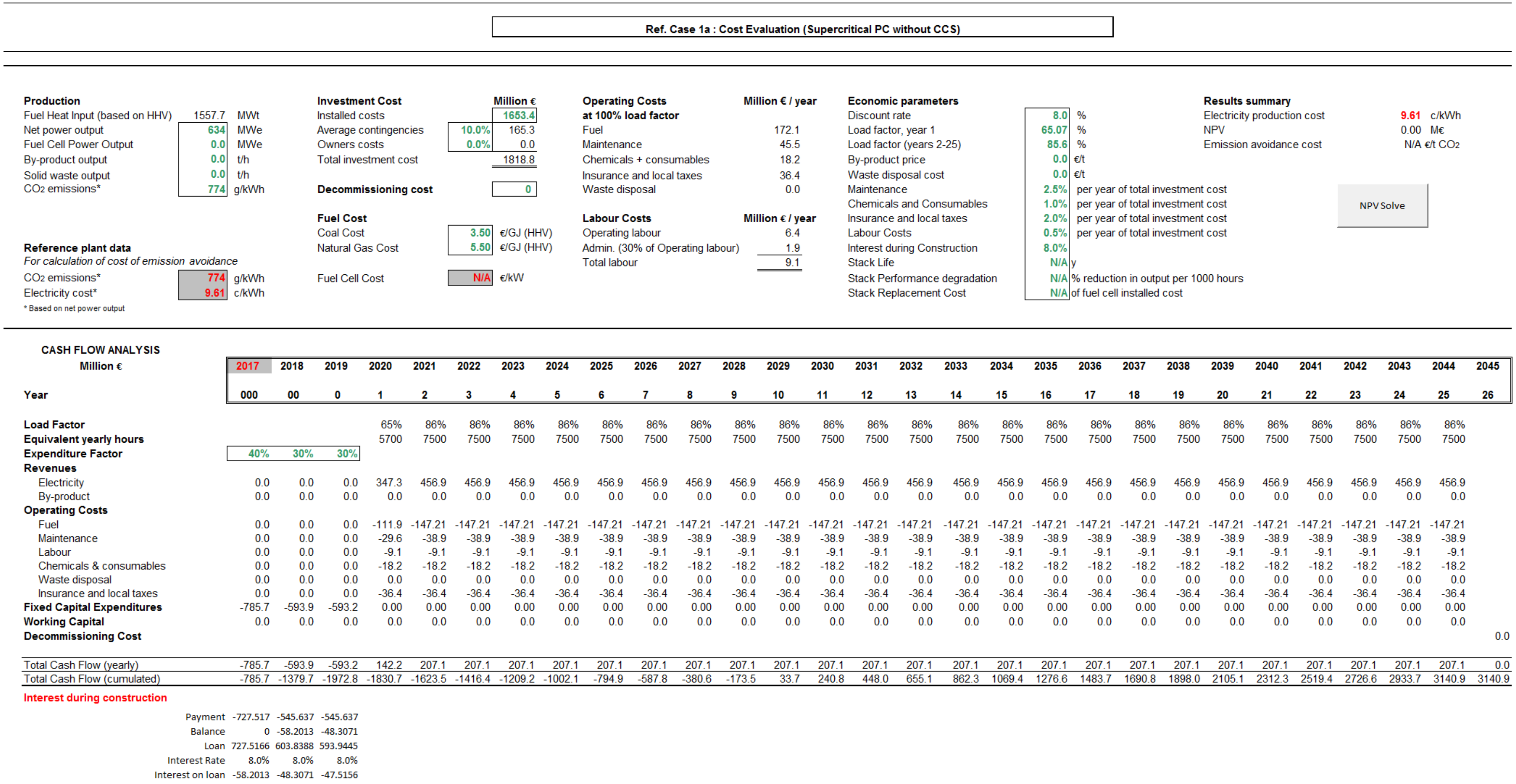
TEA Case Number	HRSG	Steam Turbine	CO2 Capture	CO2 H2O Separation	Water Gas Shift Unit	CO2 Compression	Cooling Water System	Boiler Plant and Water	Feed water and Misc. BoP
1	85.26	82.43	0.00	0.00	0.00	31.15	101.07	0.00	22.02
2	50.50	54.39	0.00	0.00	0.00	42.73	31.08	0.00	31.08
3	0.00	78.35	0.00	7.67	19.53	49.84	44.86	15.05	42.63
4	0.00	42.46	0.00	6.73	11.51	37.51	46.90	0.00	65.87
5	41.92	0.00	0.00	0.00	0.00	63.66	0.00	0.00	0.00
Ref 1	0.00	290.60	1030.98	0.00	0.00	0.00	101.53	0.00	182.91
Ref 2	71.04	119.72	607.35	0.00	0.00	0.00	44.17	0.00	93.05
Ref 3	70.47	153.71	0.00	0.00	0.00	138.80	67.00	0.00	113.95
Ref 4	0.00	272.46	0.00	0.00	0.00	0.00	71.87	0.00	152.84
Ref 5	66.96	115.67	0.00	0.00	0.00	0.00	28.38	0.00	76.81
Ref 6	64.01	137.45	0.00	0.00	0.00	0.00	54.02	0.00	95.06

Table A3-5 – Summary of All Cases Installed Costs Post-Rationalisation and Scaling (Part 3)

Basis - Netherlands, 2017, Million Euros

TEA Case Number	Condenser and Accesserories	Stack	Feed water steam generator	Accessory Electric Plant	Instrumentation and Control	Site Improvements	Building and Structures	Total Plant Cost
1	0.00	0.00	0.00	0.00	0.00	0.00	0.00	880.80
2	0.00	0.00	0.00	120.42	38.85	31.08	31.08	3481.12
3	0.00	0.00	76.20	115.42	33.35	24.82	22.60	3557.57
4	4.58	4.59	51.80	51.47	24.48	12.88	14.05	3704.03
5	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1303.58
Ref 1	0.00	73.44	0.00	152.63	51.63	29.46	116.66	3162.63
Ref 2	0.00	1.80	0.00	96.06	31.43	19.25	21.09	1329.62
Ref 3	0.00	22.64	0.00	178.02	55.18	39.02	37.25	3037.44
Ref 4	0.00	74.47	0.00	100.76	42.95	26.76	109.31	1818.79
Ref 5	0.00	1.99	0.00	65.05	24.05	16.78	19.29	614.44
Ref 6	0.00	19.29	0.00	133.65	42.84	32.39	31.54	2171.65

Table A3-5 – Summary of All Cases Installed Costs Post-Rationalisation and Scaling (Part 4)



Ref. Case 1a : Cost Evaluation (Supercritical PC without CCS)

Production		
Fuel Heat Input (based on HHV)	1557.7	MWt
Net power output	634	MWe
Fuel Cell Power Output	0.0	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	774	g/kWh

Investment Cost		Million €
Installed costs		1653.4
Average contingencies	10.0%	165.3
Owners costs	0.0%	0.0
Total investment cost		1818.8
Decommissioning cost		0

Operating Costs at 100% load factor	Million € / year
Fuel	172.1
Maintenance	45.5
Chemicals + consumables	18.2
Insurance and local taxes	36.4
Waste disposal	0.0

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	N/A	y
Stack Performance degradation	N/A	% reduction in output per 1000 hours
Stack Replacement Cost	N/A	of fuel cell installed cost

Results summary		
Electricity production cost	9.61	c/kWh
NPV	0.00	M€
Emission avoidance cost	N/A	€/t CO_2

NPV Solve

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	N/A	€/kW

Labour Costs	Million € / year
Operating labour	6.4
Admin. (30% of Operating labour)	1.9
Total labour	9.1

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	774	g/kWh
Electricity cost*	9.61	c/kWh

* Based on net power output

CASH FLOW ANALYSIS

Million €

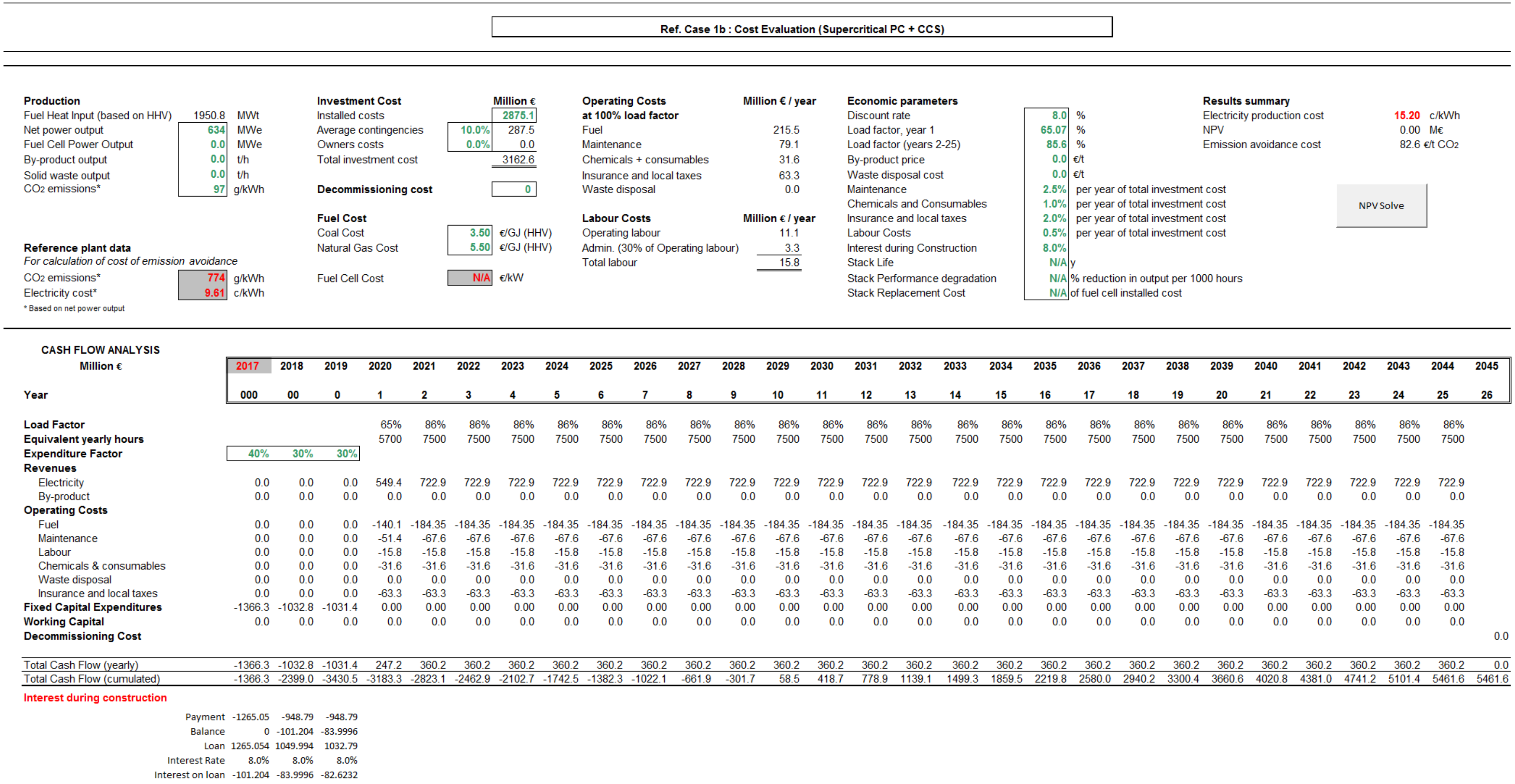
	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%												
Revenues															
Electricity	0.0	0.0	0.0	347.3	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-111.9	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21
Maintenance	0.0	0.0	0.0	-29.6	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9
Labour	0.0	0.0	0.0	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1
Chemicals & consumables	0.0	0.0	0.0	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4
Fixed Capital Expenditures	-785.7	-593.9	-593.2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-785.7	-593.9	-593.2	142.2	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1
Total Cash Flow (cumulated)	-785.7	-1379.7	-1972.8	-1830.7	-1623.5	-1416.4	-1209.2	-1002.1	-794.9	-587.8	-380.6	-173.5	33.7	240.8	448.0

	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor														
Revenues														
Electricity	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	456.9	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	-147.21	
Maintenance	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	-38.9	
Labour	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	-9.1	
Chemicals & consumables	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	-18.2	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	-36.4	
Fixed Capital Expenditures	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost														0.0
Total Cash Flow (yearly)	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	207.1	0.0
Total Cash Flow (cumulated)	655.1	862.3	1069.4	1276.6	1483.7	1690.8	1898.0	2105.1	2312.3	2519.4	2726.6	2933.7	3140.9	3140.9

Interest during construction

	2017	2018	2019
Payment	-727.517	-545.637	-545.637
Balance	0	-58.2013	-48.3071
Loan	727.5166	603.8388	593.9445
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-58.2013	-48.3071	-47.5156

Figure A3-1 – Case 1a Economic Assessment



Ref. Case 1b : Cost Evaluation (Supercritical PC + CCS)

Production		
Fuel Heat Input (based on HHV)	1950.8	MWt
Net power output	634	MWe
Fuel Cell Power Output	0.0	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	97	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	774	g/kWh
Electricity cost*	9.61	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		2875.1
Average contingencies	10.0%	287.5
Owners costs	0.0%	0.0
Total investment cost		3162.6
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	N/A	€/kW

Operating Costs **at 100% load factor**	**Million € / year**
Fuel	215.5
Maintenance	79.1
Chemicals + consumables	31.6
Insurance and local taxes	63.3
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	11.1
Admin. (30% of Operating labour)	3.3
Total labour	15.8

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	N/A	y
Stack Performance degradation	N/A	% reduction in output per 1000 hours
Stack Replacement Cost	N/A	of fuel cell installed cost

Results summary		
Electricity production cost	15.20	c/kWh
NPV	0.00	M€
Emission avoidance cost	82.6	€/t CO_2

NPV Solve

CASH FLOW ANALYSIS

Million €

	2017	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%											
Revenues														
Electricity	0.0	0.0	0.0	549.4	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-140.1	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35
Maintenance	0.0	0.0	0.0	-51.4	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6
Labour	0.0	0.0	0.0	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8
Chemicals & consumables	0.0	0.0	0.0	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3
Fixed Capital Expenditures	-1366.3	-1032.8	-1031.4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-1366.3	-1032.8	-1031.4	247.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2
Total Cash Flow (cumulated)	-1366.3	-2399.0	-3430.5	-3183.3	-2823.1	-2462.9	-2102.7	-1742.5	-1382.3	-1022.1	-661.9	-301.7	58.5	418.7

	2031	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**	**2045**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor															
Revenues															
Electricity	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	722.9	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	-184.35	
Maintenance	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	-67.6	
Labour	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	-15.8	
Chemicals & consumables	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	-31.6	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	-63.3	
Fixed Capital Expenditures	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost															0.0
Total Cash Flow (yearly)	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	360.2	0.0
Total Cash Flow (cumulated)	778.9	1139.1	1499.3	1859.5	2219.8	2580.0	2940.2	3300.4	3660.6	4020.8	4381.0	4741.2	5101.4	5461.6	5461.6

Interest during construction

Payment	-1265.05	-948.79	-948.79
Balance	0	-101.204	-83.9996
Loan	1265.054	1049.994	1032.79
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-101.204	-83.9996	-82.6232

Figure A3-2 – Case 1b Economic Assessment

Ref. Case 2a : Cost Evaluation (NGCC without CCS)

Production		
Fuel Heat Input (based on HHV)	1231.1	MWt
Net power output	634	MWe
Fuel Cell Power Output	0.0	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	357	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	357	g/kWh
Electricity cost*	6.05	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		558.6
Average contingencies	10.0%	55.9
Owners costs	0.0%	0.0
Total investment cost		614.4
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	N/A	€/kW

Operating Costs **at 100% load factor**	**Million € / year**
Fuel	213.7
Maintenance	15.4
Chemicals + consumables	6.1
Insurance and local taxes	12.3
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	2.2
Admin. (30% of Operating labour)	0.6
Total labour	3.1

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	N/A	y
Stack Performance degradation	N/A	% reduction in output per 1000 hours
Stack Replacement Cost	N/A	of fuel cell installed cost

Results summary		
Electricity production cost	6.05	c/kWh
NPV	0.00	M€
Emission avoidance cost	N/A	€/t CO_2

NPV Solve

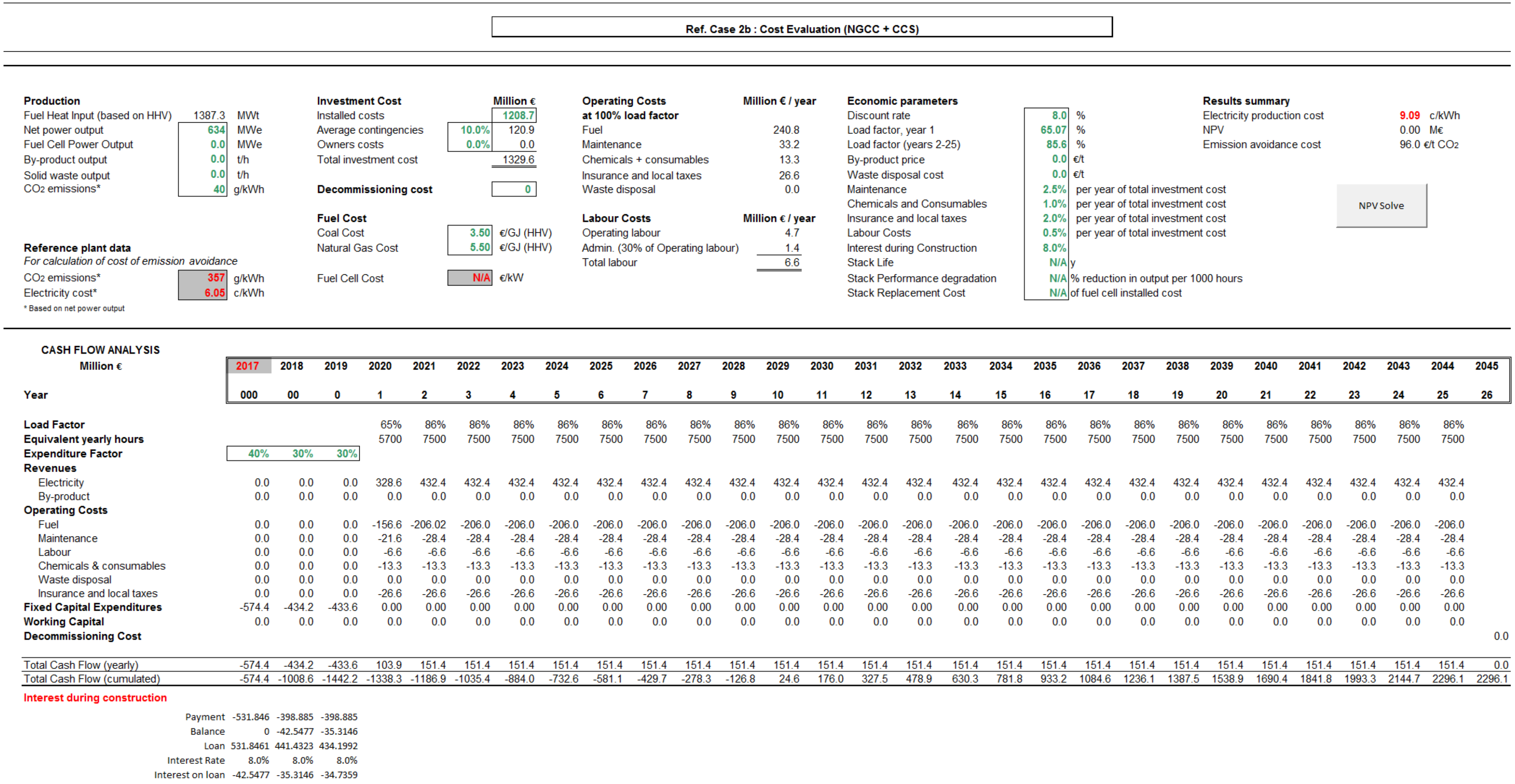
CASH FLOW ANALYSIS
Million €

	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor	40%	30%	30%																										
Revenues																													
Electricity	0.0	0.0	0.0	218.5	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	287.4	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs																													
Fuel	0.0	0.0	0.0	-138.9	-182.81	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	-182.8	
Maintenance	0.0	0.0	0.0	-10.0	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	-13.1	
Labour	0.0	0.0	0.0	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	-3.1	
Chemicals & consumables	0.0	0.0	0.0	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	-6.1	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	0.0	0.0	0.0	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	-12.3	
Fixed Capital Expenditures	-265.4	-200.7	-200.4	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost																													0.0
Total Cash Flow (yearly)	-265.4	-200.7	-200.4	48.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	70.0	0.0
Total Cash Flow (cumulated)	-265.4	-466.1	-666.5	-618.5	-548.5	-478.5	-408.5	-338.5	-268.6	-198.6	-128.6	-58.6	11.4	81.4	151.3	221.3	291.3	361.3	431.3	501.2	571.2	641.2	711.2	781.2	851.1	921.1	991.1	1061.1	1061.1

Interest during construction

Payment	-245.778	-184.333	-184.333
Balance	0	-19.6622	-16.3197
Loan	245.7779	203.9956	200.6531
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-19.6622	-16.3197	-16.0522

Figure A3-3 – Case 2a Economic Assessment



Ref. Case 2b : Cost Evaluation (NGCC + CCS)

Production		
Fuel Heat Input (based on HHV)	1387.3	MWt
Net power output	634	MWe
Fuel Cell Power Output	0.0	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO2 emissions*	40	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO2 emissions*	357	g/kWh
Electricity cost*	6.05	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		1208.7
Average contingencies	10.0%	120.9
Owners costs	0.0%	0.0
Total investment cost		1329.6
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	N/A	€/kW

Operating Costs at 100% load factor	**Million € / year**
Fuel	240.8
Maintenance	33.2
Chemicals + consumables	13.3
Insurance and local taxes	26.6
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	4.7
Admin. (30% of Operating labour)	1.4
Total labour	6.6

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	N/A	y
Stack Performance degradation	N/A	% reduction in output per 1000 hours
Stack Replacement Cost	N/A	of fuel cell installed cost

Results summary		
Electricity production cost	9.09	c/kWh
NPV	0.00	M€
Emission avoidance cost	96.0	€/t CO2

CASH FLOW ANALYSIS

Million €	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**	**2045**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor	40%	30%	30%																										
Revenues																													
Electricity	0.0	0.0	0.0	328.6	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	432.4	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs																													
Fuel	0.0	0.0	0.0	-156.6	-206.02	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	-206.0	
Maintenance	0.0	0.0	0.0	-21.6	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	-28.4	
Labour	0.0	0.0	0.0	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	-6.6	
Chemicals & consumables	0.0	0.0	0.0	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	-13.3	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	0.0	0.0	0.0	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	-26.6	
Fixed Capital Expenditures	-574.4	-434.2	-433.6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost																													0.0
Total Cash Flow (yearly)	-574.4	-434.2	-433.6	103.9	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	151.4	0.0
Total Cash Flow (cumulated)	-574.4	-1008.6	-1442.2	-1338.3	-1186.9	-1035.4	-884.0	-732.6	-581.1	-429.7	-278.3	-126.8	24.6	176.0	327.5	478.9	630.3	781.8	933.2	1084.6	1236.1	1387.5	1538.9	1690.4	1841.8	1993.3	2144.7	2296.1	2296.1

Interest during construction

Payment	-531.846	-398.885	-398.885
Balance	0	-42.5477	-35.3146
Loan	531.8461	441.4323	434.1992
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-42.5477	-35.3146	-34.7359

Figure A3-4 – Case 2b Economic Assessment

Ref. Case 3a : Cost Evaluation (IGCC without CCS)

Production		
Fuel Heat Input (based on HHV)	1625.6	MWt
Net power output	634	MWe
Fuel Cell Power Output	0.0	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	782	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	782	g/kWh
Electricity cost*	11.01	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		1974.2
Average contingencies	10.0%	197.4
Owners costs	0.0%	0.0
Total investment cost		2171.6
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	N/A	€/kW

Operating Costs **at 100% load factor**	**Million € / year**
Fuel	179.6
Maintenance	54.3
Chemicals + consumables	21.7
Insurance and local taxes	43.4
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	7.6
Admin. (30% of Operating labour)	2.3
Total labour	10.9

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	N/A	y
Stack Performance degradation	N/A	% reduction in output per 1000 hours
Stack Replacement Cost	N/A	of fuel cell installed cost

Results summary		
Electricity production cost	11.01	c/kWh
NPV	0.00	M€
Emission avoidance cost	N/A	€/t CO_2

NPV Solve

CASH FLOW ANALYSIS
Million €

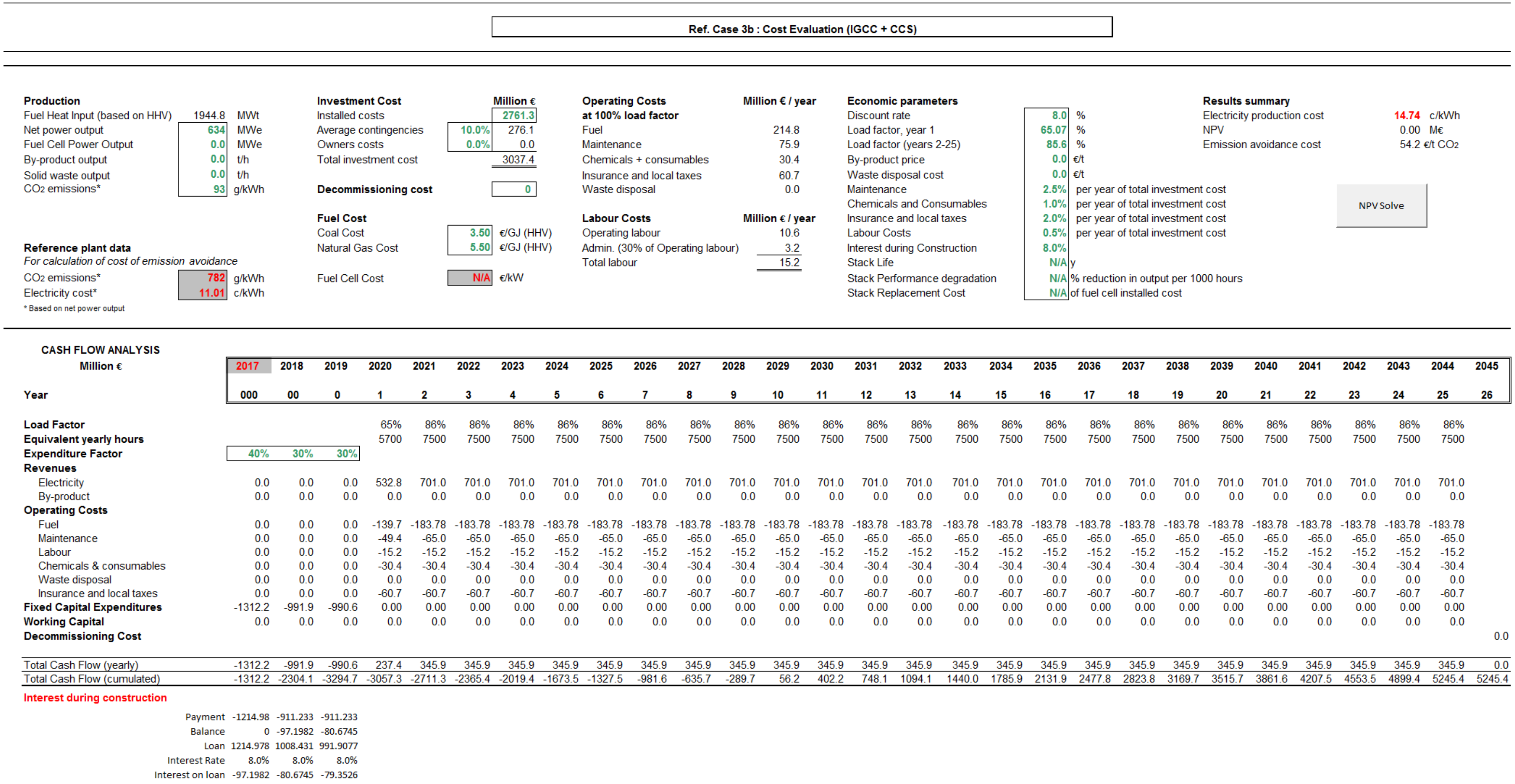
	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%												
Revenues															
Electricity	0.0	0.0	0.0	397.8	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-116.8	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62
Maintenance	0.0	0.0	0.0	-35.3	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5
Labour	0.0	0.0	0.0	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9
Chemicals & consumables	0.0	0.0	0.0	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4
Fixed Capital Expenditures	-938.2	-709.2	-708.2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-938.2	-709.2	-708.2	169.7	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3
Total Cash Flow (cumulated)	-938.2	-1647.3	-2355.6	-2185.8	-1938.5	-1691.2	-1443.8	-1196.5	-949.1	-701.8	-454.5	-207.1	40.2	287.5	534.9

	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor														
Revenues														
Electricity	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	523.4	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	-153.62	
Maintenance	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	-46.5	
Labour	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	
Chemicals & consumables	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	-21.7	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	-43.4	
Fixed Capital Expenditures	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost														0.0
Total Cash Flow (yearly)	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	247.3	0.0
Total Cash Flow (cumulated)	782.2	1029.5	1276.9	1524.2	1771.5	2018.9	2266.2	2513.6	2760.9	3008.2	3255.6	3502.9	3750.2	3750.2

Interest during construction

Payment	-868.659	-651.495	-651.495
Balance	0	-69.4927	-57.679
Loan	868.6594	720.9873	709.1735
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-69.4927	-57.679	-56.7339

Figure A3-5 – Case 3a Economic Assessment



Ref. Case 3b : Cost Evaluation (IGCC + CCS)

Production		
Fuel Heat Input (based on HHV)	1944.8	MWt
Net power output	634	MWe
Fuel Cell Power Output	0.0	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	93	g/kWh

Investment Cost		**Million €**
Installed costs		2761.3
Average contingencies	10.0%	276.1
Owners costs	0.0%	0.0
Total investment cost		3037.4
Decommissioning cost		0

Operating Costs	**Million € / year**
at 100% load factor	
Fuel	214.8
Maintenance	75.9
Chemicals + consumables	30.4
Insurance and local taxes	60.7
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	10.6
Admin. (30% of Operating labour)	3.2
Total labour	15.2

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	N/A	€/kW

Reference plant data		
For calculation of cost of emission avoidance		
CO_2 emissions*	782	g/kWh
Electricity cost*	11.01	c/kWh

* Based on net power output

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	N/A	y
Stack Performance degradation	N/A	% reduction in output per 1000 hours
Stack Replacement Cost	N/A	of fuel cell installed cost

Results summary		
Electricity production cost	14.74	c/kWh
NPV	0.00	M€
Emission avoidance cost	54.2	€/t CO_2

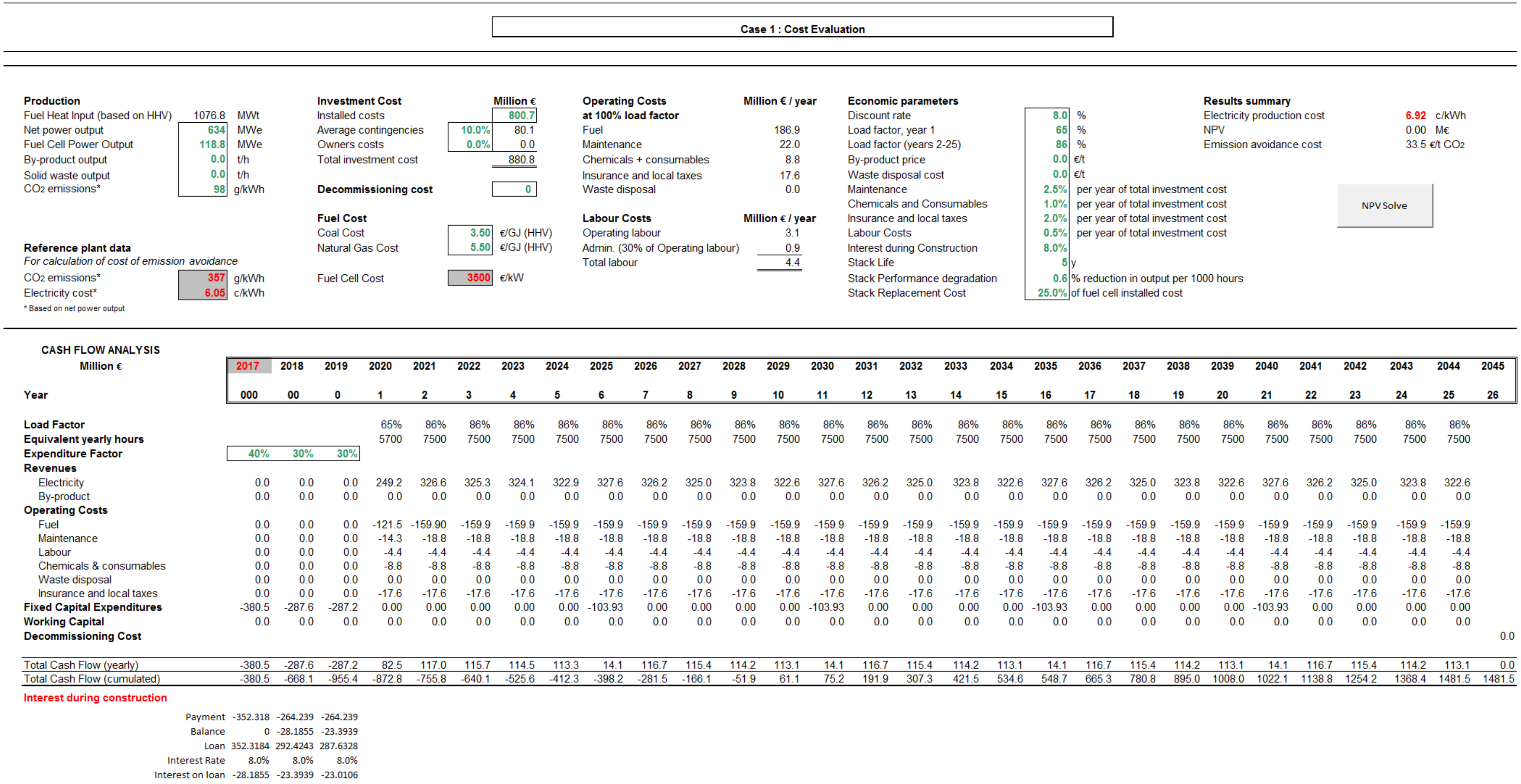
NPV Solve

CASH FLOW ANALYSIS **Million €**	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor	40%	30%	30%																										
Revenues																													
Electricity	0.0	0.0	0.0	532.8	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	701.0	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs																													
Fuel	0.0	0.0	0.0	-139.7	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	-183.78	
Maintenance	0.0	0.0	0.0	-49.4	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	-65.0	
Labour	0.0	0.0	0.0	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	-15.2	
Chemicals & consumables	0.0	0.0	0.0	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	-30.4	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	0.0	0.0	0.0	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	-60.7	
Fixed Capital Expenditures	-1312.2	-991.9	-990.6	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost																													0.0
Total Cash Flow (yearly)	-1312.2	-991.9	-990.6	237.4	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	345.9	0.0
Total Cash Flow (cumulated)	-1312.2	-2304.1	-3294.7	-3057.3	-2711.3	-2365.4	-2019.4	-1673.5	-1327.5	-981.6	-635.7	-289.7	56.2	402.2	748.1	1094.1	1440.0	1785.9	2131.9	2477.8	2823.8	3169.7	3515.7	3861.6	4207.5	4553.5	4899.4	5245.4	5245.4

Interest during construction

Payment	-1214.98	-911.233	-911.233
Balance	0	-97.1982	-80.6745
Loan	1214.978	1008.431	991.9077
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-97.1982	-80.6745	-79.3526

Figure A3-6 – Case 3b Economic Assessment



Case 1 : Cost Evaluation

Production		
Fuel Heat Input (based on HHV)	1076.8	MWt
Net power output	634	MWe
Fuel Cell Power Output	118.8	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	98	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	357	g/kWh
Electricity cost*	6.05	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		800.7
Average contingencies	10.0%	80.1
Owners costs	0.0%	0.0
Total investment cost		880.8
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	3500	€/kW

Operating Costs **at 100% load factor**	**Million € / year**
Fuel	186.9
Maintenance	22.0
Chemicals + consumables	8.8
Insurance and local taxes	17.6
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	3.1
Admin. (30% of Operating labour)	0.9
Total labour	4.4

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65	%
Load factor (years 2-25)	86	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	5	y
Stack Performance degradation	0.6	% reduction in output per 1000 hours
Stack Replacement Cost	25.0%	of fuel cell installed cost

Results summary		
Electricity production cost	6.92	c/kWh
NPV	0.00	M€
Emission avoidance cost	33.5	€/t CO_2

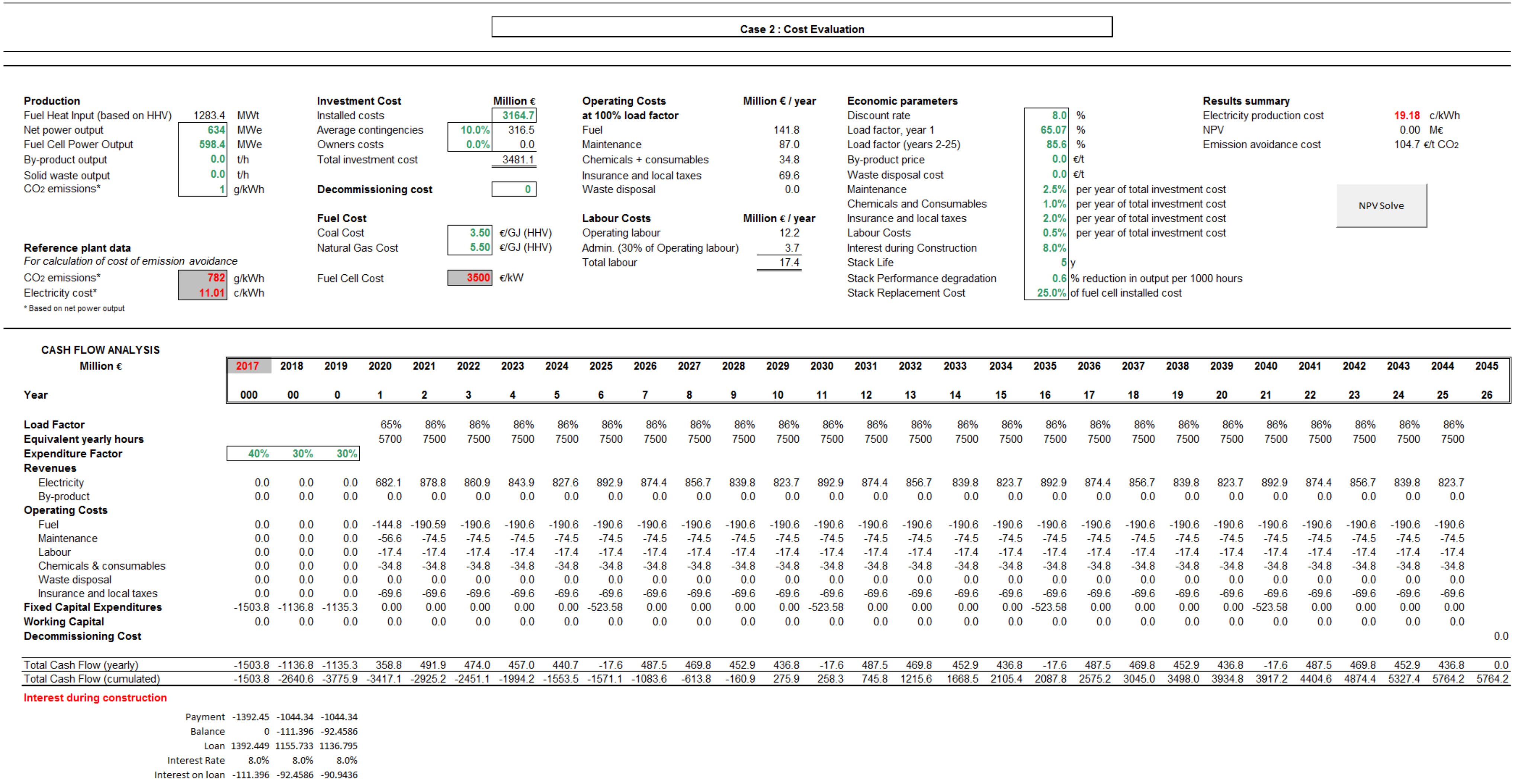
CASH FLOW ANALYSIS **Million €**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%												
Revenues															
Electricity	0.0	0.0	0.0	249.2	326.6	325.3	324.1	322.9	327.6	326.2	325.0	323.8	322.6	327.6	326.2
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-121.5	-159.90	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9
Maintenance	0.0	0.0	0.0	-14.3	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8
Labour	0.0	0.0	0.0	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4
Chemicals & consumables	0.0	0.0	0.0	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6
Fixed Capital Expenditures	-380.5	-287.6	-287.2	0.00	0.00	0.00	0.00	0.00	-103.93	0.00	0.00	0.00	0.00	-103.93	0.00
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-380.5	-287.6	-287.2	82.5	117.0	115.7	114.5	113.3	14.1	116.7	115.4	114.2	113.1	14.1	116.7
Total Cash Flow (cumulated)	-380.5	-668.1	-955.4	-872.8	-755.8	-640.1	-525.6	-412.3	-398.2	-281.5	-166.1	-51.9	61.1	75.2	191.9

CASH FLOW ANALYSIS **Million €**	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**	**2045**
Year	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor														
Revenues														
Electricity	325.0	323.8	322.6	327.6	326.2	325.0	323.8	322.6	327.6	326.2	325.0	323.8	322.6	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	-159.9	
Maintenance	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	-18.8	
Labour	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	-4.4	
Chemicals & consumables	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	-8.8	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	-17.6	
Fixed Capital Expenditures	0.00	0.00	0.00	-103.93	0.00	0.00	0.00	0.00	-103.93	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost														0.0
Total Cash Flow (yearly)	115.4	114.2	113.1	14.1	116.7	115.4	114.2	113.1	14.1	116.7	115.4	114.2	113.1	0.0
Total Cash Flow (cumulated)	307.3	421.5	534.6	548.7	665.3	780.8	895.0	1008.0	1022.1	1138.8	1254.2	1368.4	1481.5	1481.5

Interest during construction

Payment	-352.318	-264.239	-264.239
Balance	0	-28.1855	-23.3939
Loan	352.3184	292.4243	287.6328
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-28.1855	-23.3939	-23.0106

Figure A3-7 – Case 1 Economic Assessment



Case 2 : Cost Evaluation

Production		
Fuel Heat Input (based on HHV)	1283.4	MWt
Net power output	634	MWe
Fuel Cell Power Output	598.4	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	1	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	782	g/kWh
Electricity cost*	11.01	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		3164.7
Average contingencies	10.0%	316.5
Owners costs	0.0%	0.0
Total investment cost		3481.1
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	3500	€/kW

Operating Costs **at 100% load factor**	**Million € / year**
Fuel	141.8
Maintenance	87.0
Chemicals + consumables	34.8
Insurance and local taxes	69.6
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	12.2
Admin. (30% of Operating labour)	3.7
Total labour	17.4

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	5	y
Stack Performance degradation	0.6	% reduction in output per 1000 hours
Stack Replacement Cost	25.0%	of fuel cell installed cost

Results summary		
Electricity production cost	19.18	c/kWh
NPV	0.00	M€
Emission avoidance cost	104.7	€/t CO_2

NPV Solve

CASH FLOW ANALYSIS

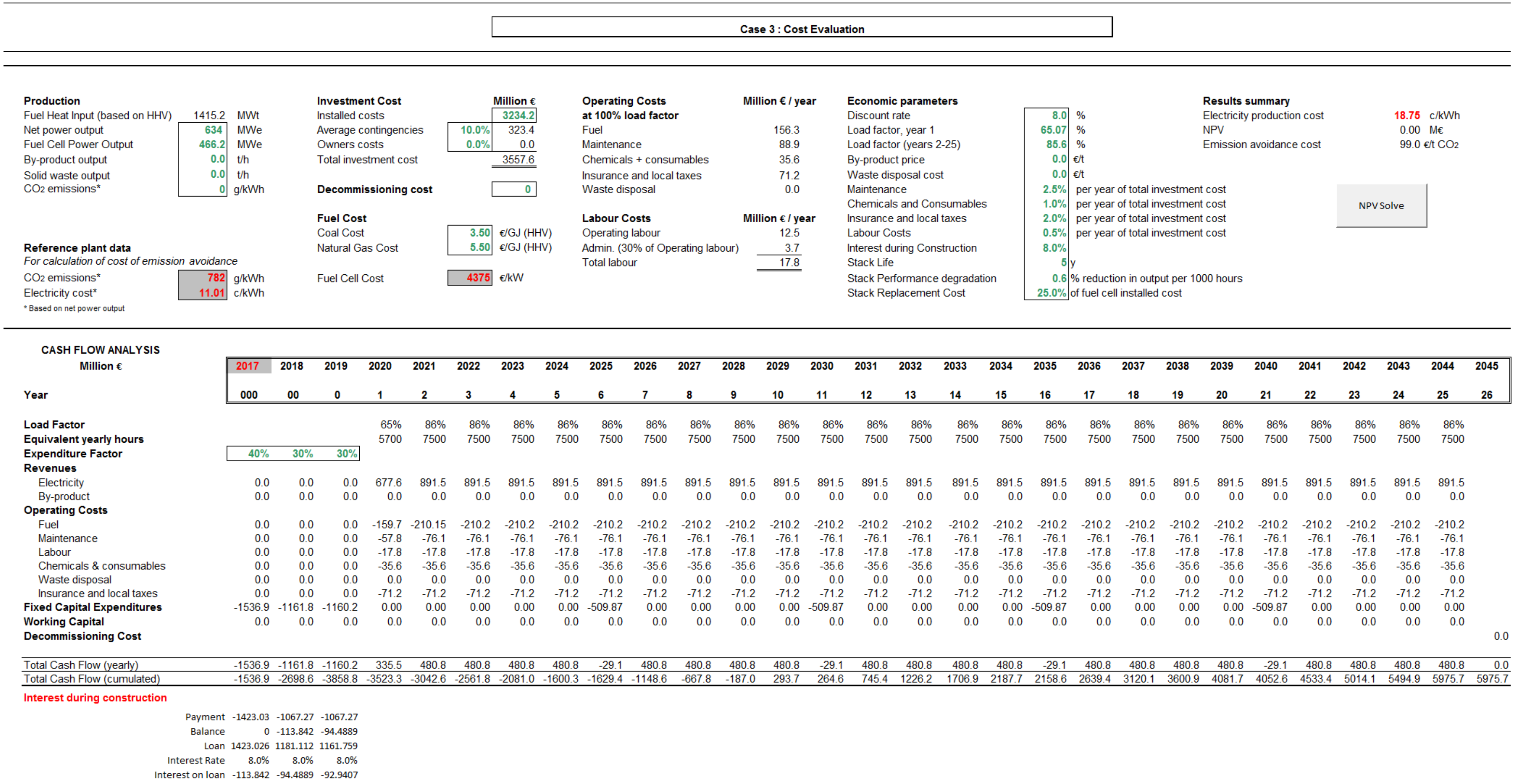
Million €	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%											
Revenues														
Electricity	0.0	0.0	0.0	682.1	878.8	860.9	843.9	827.6	892.9	874.4	856.7	839.8	823.7	892.9
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-144.8	-190.59	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6
Maintenance	0.0	0.0	0.0	-56.6	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5
Labour	0.0	0.0	0.0	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4
Chemicals & consumables	0.0	0.0	0.0	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6
Fixed Capital Expenditures	-1503.8	-1136.8	-1135.3	0.00	0.00	0.00	0.00	0.00	-523.58	0.00	0.00	0.00	0.00	-523.58
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-1503.8	-1136.8	-1135.3	358.8	491.9	474.0	457.0	440.7	-17.6	487.5	469.8	452.9	436.8	-17.6
Total Cash Flow (cumulated)	-1503.8	-2640.6	-3775.9	-3417.1	-2925.2	-2451.1	-1994.2	-1553.5	-1571.1	-1083.6	-613.8	-160.9	275.9	258.3

Million €	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**	**2045**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor															
Revenues															
Electricity	874.4	856.7	839.8	823.7	892.9	874.4	856.7	839.8	823.7	892.9	874.4	856.7	839.8	823.7	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	-190.6	
Maintenance	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	-74.5	
Labour	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	-17.4	
Chemicals & consumables	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	-34.8	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	-69.6	
Fixed Capital Expenditures	0.00	0.00	0.00	0.00	-523.58	0.00	0.00	0.00	0.00	-523.58	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost															0.0
Total Cash Flow (yearly)	487.5	469.8	452.9	436.8	-17.6	487.5	469.8	452.9	436.8	-17.6	487.5	469.8	452.9	436.8	0.0
Total Cash Flow (cumulated)	745.8	1215.6	1668.5	2105.4	2087.8	2575.2	3045.0	3498.0	3934.8	3917.2	4404.6	4874.4	5327.4	5764.2	5764.2

Interest during construction

Payment	-1392.45	-1044.34	-1044.34
Balance	0	-111.396	-92.4586
Loan	1392.449	1155.733	1136.795
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-111.396	-92.4586	-90.9436

Figure A3-8 – Case 2 Economic Assessment



Case 3 : Cost Evaluation

Production		
Fuel Heat Input (based on HHV)	1415.2	MWt
Net power output	634	MWe
Fuel Cell Power Output	466.2	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	0	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	782	g/kWh
Electricity cost*	11.01	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		3234.2
Average contingencies	10.0%	323.4
Owners costs	0.0%	0.0
Total investment cost		3557.6
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	4375	€/kW

Operating Costs **at 100% load factor**	**Million € / year**
Fuel	156.3
Maintenance	88.9
Chemicals + consumables	35.6
Insurance and local taxes	71.2
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	12.5
Admin. (30% of Operating labour)	3.7
Total labour	17.8

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	5	y
Stack Performance degradation	0.6	% reduction in output per 1000 hours
Stack Replacement Cost	25.0%	of fuel cell installed cost

Results summary		
Electricity production cost	18.75	c/kWh
NPV	0.00	M€
Emission avoidance cost	99.0	€/t CO_2

NPV Solve

CASH FLOW ANALYSIS
Million €

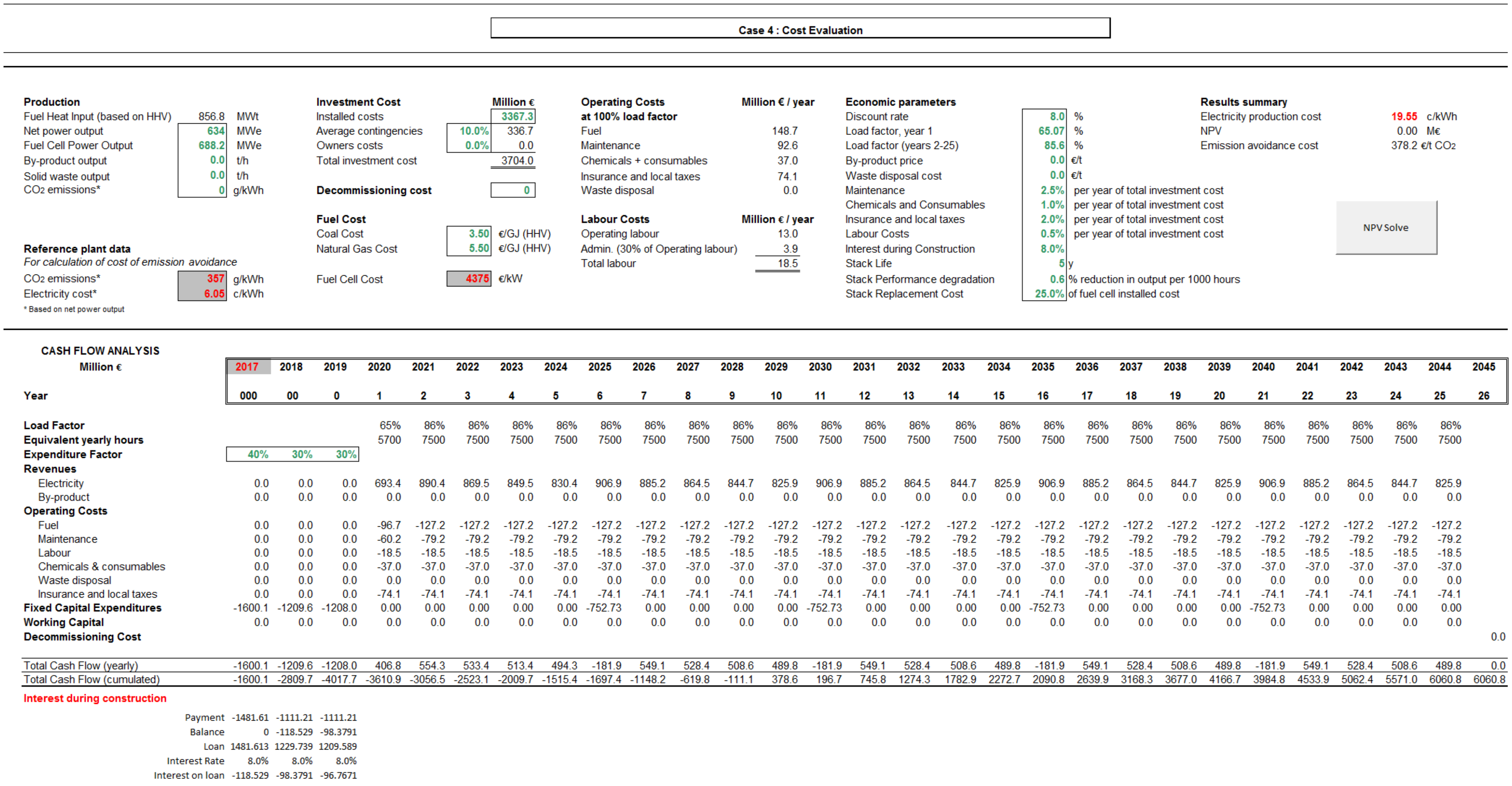
	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%											
Revenues														
Electricity	0.0	0.0	0.0	677.6	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-159.7	-210.15	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2
Maintenance	0.0	0.0	0.0	-57.8	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1
Labour	0.0	0.0	0.0	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8
Chemicals & consumables	0.0	0.0	0.0	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2
Fixed Capital Expenditures	-1536.9	-1161.8	-1160.2	0.00	0.00	0.00	0.00	0.00	-509.87	0.00	0.00	0.00	0.00	-509.87
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-1536.9	-1161.8	-1160.2	335.5	480.8	480.8	480.8	480.8	-29.1	480.8	480.8	480.8	480.8	-29.1
Total Cash Flow (cumulated)	-1536.9	-2698.6	-3858.8	-3523.3	-3042.6	-2561.8	-2081.0	-1600.3	-1629.4	-1148.6	-667.8	-187.0	293.7	264.6

	2031	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor															
Revenues															
Electricity	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	891.5	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	-210.2	
Maintenance	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	-76.1	
Labour	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	-17.8	
Chemicals & consumables	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	-35.6	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	-71.2	
Fixed Capital Expenditures	0.00	0.00	0.00	0.00	-509.87	0.00	0.00	0.00	0.00	-509.87	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost															0.0
Total Cash Flow (yearly)	480.8	480.8	480.8	480.8	-29.1	480.8	480.8	480.8	480.8	-29.1	480.8	480.8	480.8	480.8	0.0
Total Cash Flow (cumulated)	745.4	1226.2	1706.9	2187.7	2158.6	2639.4	3120.1	3600.9	4081.7	4052.6	4533.4	5014.1	5494.9	5975.7	5975.7

Interest during construction

Payment	-1423.03	-1067.27	-1067.27
Balance	0	-113.842	-94.4889
Loan	1423.026	1181.112	1161.759
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-113.842	-94.4889	-92.9407

Figure A3-9 – Case 3 Economic Assessment



Case 4 : Cost Evaluation

Production		
Fuel Heat Input (based on HHV)	856.8	MWt
Net power output	634	MWe
Fuel Cell Power Output	688.2	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO2 emissions*	0	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO2 emissions*	357	g/kWh
Electricity cost*	6.05	c/kWh

* Based on net power output

Investment Cost		**Million €**
Installed costs		3367.3
Average contingencies	10.0%	336.7
Owners costs	0.0%	0.0
Total investment cost		3704.0
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	4375	€/kW

Operating Costs	**Million € / year**
at 100% load factor	
Fuel	148.7
Maintenance	92.6
Chemicals + consumables	37.0
Insurance and local taxes	74.1
Waste disposal	0.0

Labour Costs	**Million € / year**
Operating labour	13.0
Admin. (30% of Operating labour)	3.9
Total labour	18.5

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	5	y
Stack Performance degradation	0.6	% reduction in output per 1000 hours
Stack Replacement Cost	25.0%	of fuel cell installed cost

Results summary		
Electricity production cost	19.55	c/kWh
NPV	0.00	M€
Emission avoidance cost	378.2	€/t CO2

CASH FLOW ANALYSIS **Million €**	**2017**	**2018**	**2019**	**2020**	**2021**	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%											
Revenues														
Electricity	0.0	0.0	0.0	693.4	890.4	869.5	849.5	830.4	906.9	885.2	864.5	844.7	825.9	906.9
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	-96.7	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2
Maintenance	0.0	0.0	0.0	-60.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2
Labour	0.0	0.0	0.0	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5
Chemicals & consumables	0.0	0.0	0.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1
Fixed Capital Expenditures	-1600.1	-1209.6	-1208.0	0.00	0.00	0.00	0.00	0.00	-752.73	0.00	0.00	0.00	0.00	-752.73
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost														
Total Cash Flow (yearly)	-1600.1	-1209.6	-1208.0	406.8	554.3	533.4	513.4	494.3	-181.9	549.1	528.4	508.6	489.8	-181.9
Total Cash Flow (cumulated)	-1600.1	-2809.7	-4017.7	-3610.9	-3056.5	-2523.1	-2009.7	-1515.4	-1697.4	-1148.2	-619.8	-111.1	378.6	196.7

CASH FLOW ANALYSIS **Million €**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**	**2037**	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**	**2045**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor															
Revenues															
Electricity	885.2	864.5	844.7	825.9	906.9	885.2	864.5	844.7	825.9	906.9	885.2	864.5	844.7	825.9	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs															
Fuel	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	-127.2	
Maintenance	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	-79.2	
Labour	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	-18.5	
Chemicals & consumables	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	-37.0	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	-74.1	
Fixed Capital Expenditures	0.00	0.00	0.00	0.00	-752.73	0.00	0.00	0.00	0.00	-752.73	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost															0.0
Total Cash Flow (yearly)	549.1	528.4	508.6	489.8	-181.9	549.1	528.4	508.6	489.8	-181.9	549.1	528.4	508.6	489.8	0.0
Total Cash Flow (cumulated)	745.8	1274.3	1782.9	2272.7	2090.8	2639.9	3168.3	3677.0	4166.7	3984.8	4533.9	5062.4	5571.0	6060.8	6060.8

Interest during construction

Payment	-1481.61	-1111.21	-1111.21
Balance	0	-118.529	-98.3791
Loan	1481.613	1229.739	1209.589
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-118.529	-98.3791	-96.7671

Figure A3-10 – Case 4 Economic Assessment

Case 5 : Cost Evaluation

Production		
Fuel Heat Input (based on HHV)	1389.7	MWt
Net power output	634	MWe
Fuel Cell Power Output	212.8	MWe
By-product output	0.0	t/h
Solid waste output	0.0	t/h
CO_2 emissions*	113	g/kWh

Reference plant data

For calculation of cost of emission avoidance

CO_2 emissions*	774	g/kWh
Electricity cost*	9.61	c/kWh

* Based on net power output

Investment Cost		Million €
Installed costs		1185.1
Average contingencies	10.0%	118.5
Owners costs	0.0%	0.0
Total investment cost		1303.6
Decommissioning cost		0

Fuel Cost		
Coal Cost	3.50	€/GJ (HHV)
Natural Gas Cost	5.50	€/GJ (HHV)
Fuel Cell Cost	3500	€/kW

Operating Costs at 100% load factor	Million € / year
Fuel	177.7
Maintenance	32.6
Chemicals + consumables	13.0
Insurance and local taxes	26.1
Waste disposal	0.0

Labour Costs	Million € / year
Operating labour	4.6
Admin. (30% of Operating labour)	1.4
Total labour	6.5

Economic parameters		
Discount rate	8.0	%
Load factor, year 1	65.07	%
Load factor (years 2-25)	85.6	%
By-product price	0.0	€/t
Waste disposal cost	0.0	€/t
Maintenance	2.5%	per year of total investment cost
Chemicals and Consumables	1.0%	per year of total investment cost
Insurance and local taxes	2.0%	per year of total investment cost
Labour Costs	0.5%	per year of total investment cost
Interest during Construction	8.0%	
Stack Life	5	y
Stack Performance degradation	0.6	% reduction in output per 1000 hours
Stack Replacement Cost	25.0%	of fuel cell installed cost

Results summary		
Electricity production cost	8.62	c/kWh
NPV	0.00	M€
Emission avoidance cost	-15.0	€/t CO_2

NPV Solve

CASH FLOW ANALYSIS

Million €	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				65%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%
Equivalent yearly hours				5700	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500
Expenditure Factor	40%	30%	30%												
Revenues															
Electricity	0.0	0.0	0.0	309.7	404.5	401.6	398.9	396.3	406.7	403.8	400.9	398.2	395.7	406.7	403.8
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	-115.6	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0
Maintenance	0.0	0.0	0.0	-21.2	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9
Labour	0.0	0.0	0.0	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5
Chemicals & consumables	0.0	0.0	0.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Insurance and local taxes	0.0	0.0	0.0	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1
Fixed Capital Expenditures	-563.1	-425.7	-425.1	0.00	0.00	0.00	0.00	0.00	-186.18	0.00	0.00	0.00	0.00	-186.18	0.00
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost															
Total Cash Flow (yearly)	-563.1	-425.7	-425.1	127.3	178.9	176.1	173.3	170.7	-5.0	178.2	175.4	172.7	170.1	-5.0	178.2
Total Cash Flow (cumulated)	-563.1	-988.8	-1414.0	-1286.7	-1107.8	-931.7	-758.4	-587.6	-592.7	-414.4	-239.1	-66.4	103.7	98.7	276.9

Million €	2032	2033	2034	2035	2036	2037	2038	2039	2040	2041	2042	2043	2044	2045
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	86%	
Equivalent yearly hours	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	7500	
Expenditure Factor														
Revenues														
Electricity	400.9	398.2	395.7	406.7	403.8	400.9	398.2	395.7	406.7	403.8	400.9	398.2	395.7	
By-product	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Operating Costs														
Fuel	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	-152.0	
Maintenance	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	-27.9	
Labour	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	-6.5	
Chemicals & consumables	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	-13.0	
Waste disposal	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Insurance and local taxes	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	-26.1	
Fixed Capital Expenditures	0.00	0.00	0.00	-186.18	0.00	0.00	0.00	0.00	-186.18	0.00	0.00	0.00	0.00	
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	
Decommissioning Cost														0.0
Total Cash Flow (yearly)	175.4	172.7	170.1	-5.0	178.2	175.4	172.7	170.1	-5.0	178.2	175.4	172.7	170.1	0.0
Total Cash Flow (cumulated)	452.3	625.0	795.1	790.1	968.3	1143.7	1316.4	1486.5	1481.5	1659.7	1835.1	2007.8	2177.9	2177.9

Interest during construction

Payment	-521.433	-391.075	-391.075
Balance	0	-41.7147	-34.6232
Loan	521.4334	432.7897	425.6982
Interest Rate	8.0%	8.0%	8.0%
Interest on loan	-41.7147	-34.6232	-34.0559

Figure A3-11 – Case 5 Economic Assessment

IEA Greenhouse Gas R&D Programme

Pure Offices, Cheltenham Office Park, Hatherley Lane,
Cheltenham, Glos. GL51 6SH, UK

Tel: +44 1242 802911
mail@ieaghg.org
www.ieaghg.org