Project Remo2val

End of Phase Meeting

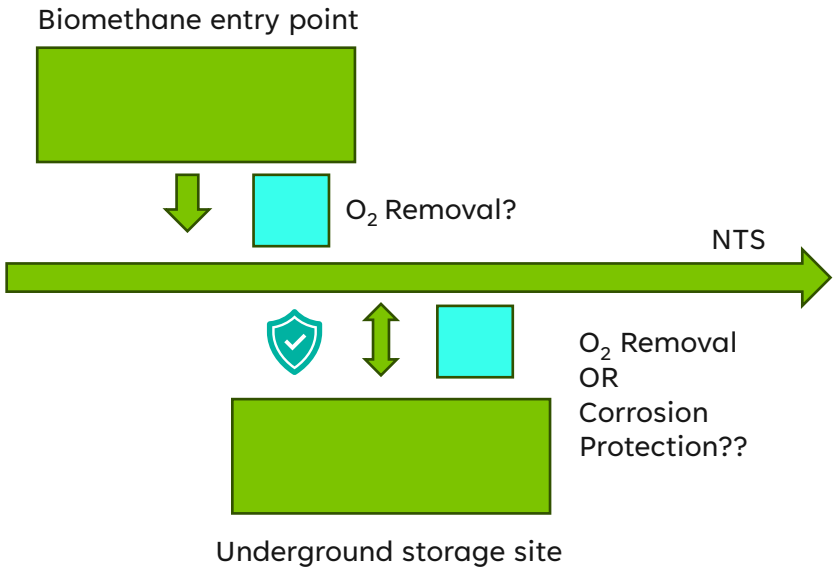
22.01.26

Project Background

Overview:

At National Gas we plan to ask for an exemption that would allow NGT and parts of the DNs that operate above 38bar to legally convey gas with up to **1mol% O_2 content** subject to following a methodology assessing the impacts. The aim is for the NTS to **enable more biomethane** onto the network. However, storage operators have raised concerns around increased **corrosion risk in wet gas conditions** to their pipelines with elevated O_2 levels >10ppm.

Project RemO2val is an innovative project aimed at **removing oxygen** from biomethane or **utilising inhibitors/coatings** to **prevent corrosion** in wet gas conditions within underground storage sites. The project will leverage advanced catalytic or adsorption technologies to ensure the integrity and longevity of storage infrastructure. The project is a **feasibility study** to understand the options and consider a CBA prior to progressing to more detailed design or potentially technical demonstrator phases



Objectives:

1. **Prevent Corrosion:** Eliminate oxygen from biomethane to acceptable levels or utilise inhibitors/coatings to prevent corrosion in wet gas environments, ensuring the durability of underground storage sites.
2. **Enhance Storage Efficiency:** Improve the quality of stored biomethane, making it safer and more efficient for long-term storage and transportation.
3. **Sustainability:** Utilise environmentally friendly technologies to achieve oxygen removal, supporting the transition to low carbon gases.
4. **Cost Efficiency**: Reduce maintenance and replacement costs associated with corrosion damage in storage facilities.
5. **Commerciality**: Mitigate risks of corrosion at storage sites to enable 1% exemption on O_2 levels so more volumes of biomethane can connect to the NTS

Project Outputs Summary

- Unable to find a market ready **corrosion inhibitor** for oxygen related internal corrosion for the proposed application in UK.
- US **'ChampionX'** oxygen inhibitor available on market but does not have **regulatory approval** for use in the UK.
- **Alternative Option: Internal lining**
- Study concluded that in-situ internal lining of pipelines likely to be **expensive** when compared to corrosion inhibitors and **not feasible** for certain sections of storage from **wellhead to cavern in subsurface.**
- Cost estimate to deploy **oxygen removal** technology at biomethane entry point **~£1M** per unit when considering a **+/- 20%** uncertainty factor.
- Cost estimate for internal lining of pipelines likely to vary between **~£1.2M - 4.2M** for **3-5km** at diameters **300-750mm**. Costs are variable and highly dependant on specific site conditions.
- **Uncertainty** as to whether we need to deploy **oxygen removal** at every **new/existing biomethane entry point** or only near **sensitive customers** e.g. storage sites/power stations/interconnectors.
- Due to a **technical challenge**, **oxygen removal** technology can only be deployed at **biomethane entry points** due to **low oxygen concentrations and high flow rates** at storage sites.
- **Conflict on interest** in terms of **needing an AGI** with **security fence, telemetry kiosk and oxygen removal deployment** identified with MOC in a pit standardised biomethane connection project.
- Cost to deploy oxygen removal at every biomethane entry point **~£1.25B** to reach **120TWh** potential by 2050.
- Cost to linepipe **~£41M** for all storage sites but certain sections of line will **not be feasible** to line e.g. cavern to wellhead.

WP1 Project Management

Tasks completed:

- Weekly PM meetings held covering actions register, decision log and project governance with minutes taken.
- Contract signed and PO's issued

To do:

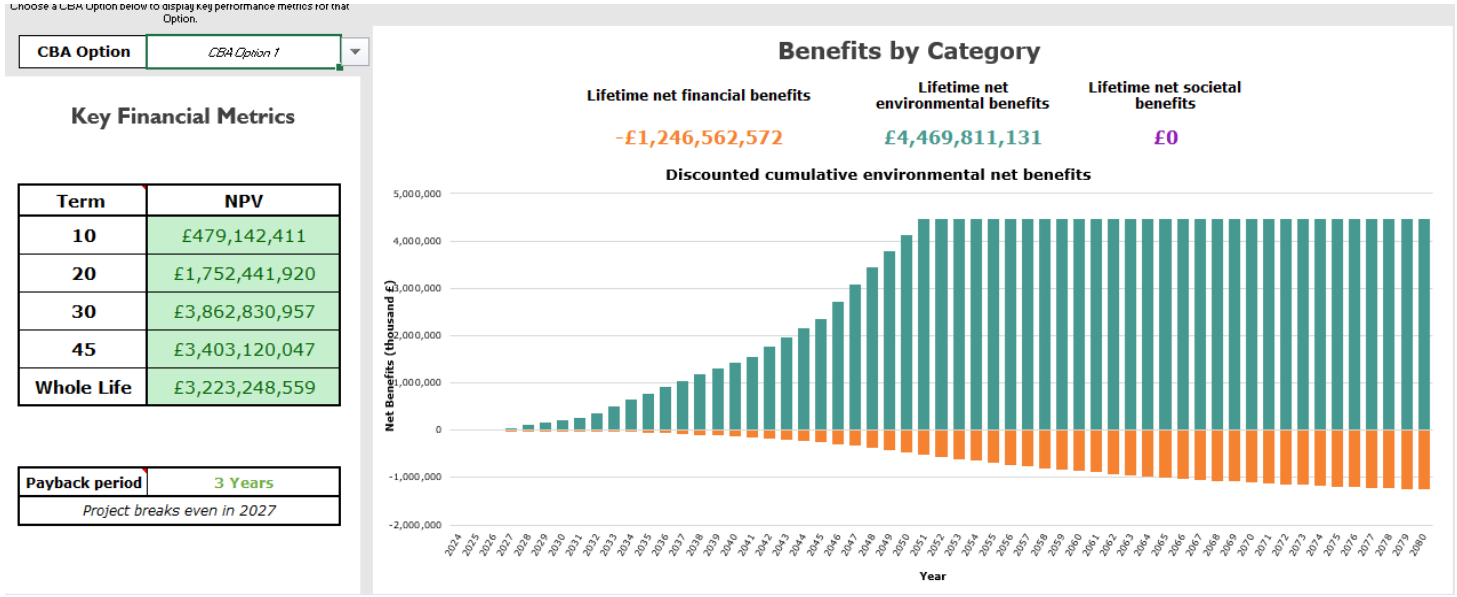
Show and tell to organise

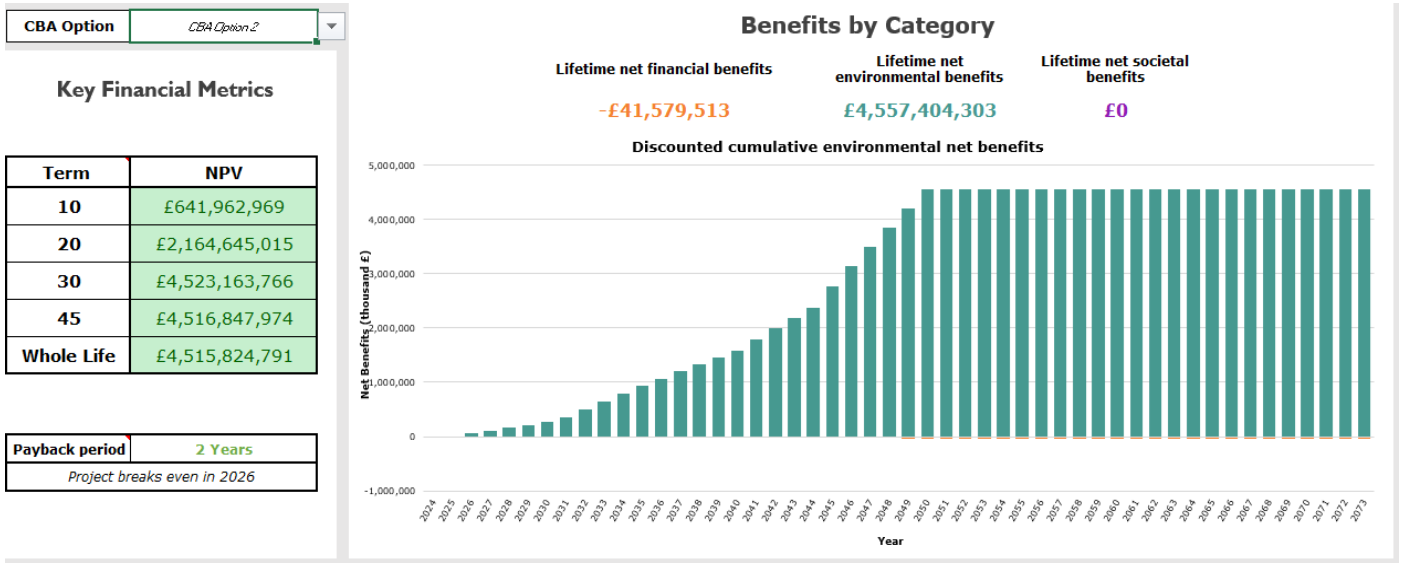
Engagement and dissemination of findings with biomethane producers via biomethane forum.

WP1 Business Case

- CBA shows that the cost to deploy oxygen removal at every entry point(**2250 sites**) would be **~£1.25B** GBP to reach **120TWh of biomethane** by 2050. If the oxygen removal only needs to be deployed around **sensitive customers**, then these costs can be reduced.

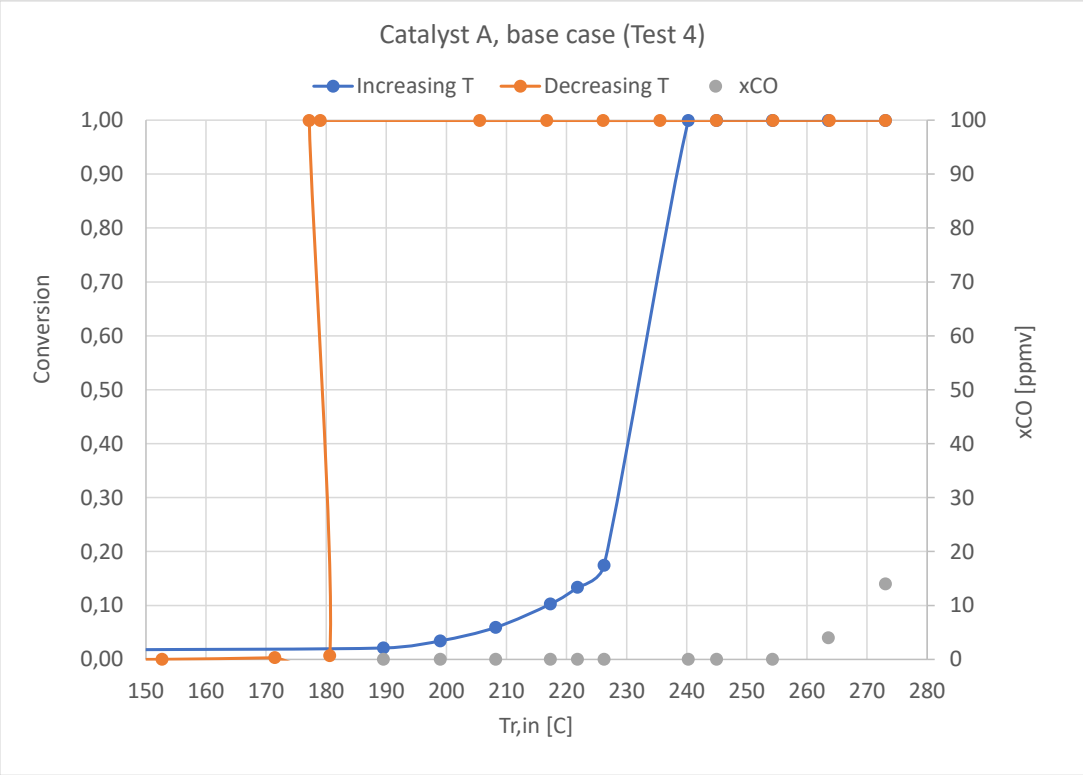
- Costs to line the pipe at every storage site (8 existing sites) is **~£41M**. However, this may go up slightly once we get exact lengths and diameters of wet gas pipework needing to be lined from storage operators for all sites. Also, it is not feasible to line the pipe from cavern to wellhead putting the completion tubing of the salt caverns at high risk of corrosion. **Option deemed not technically feasible for all wet gas pipework lines.**





WP2 Oxygen Removal

Results catalyst testing

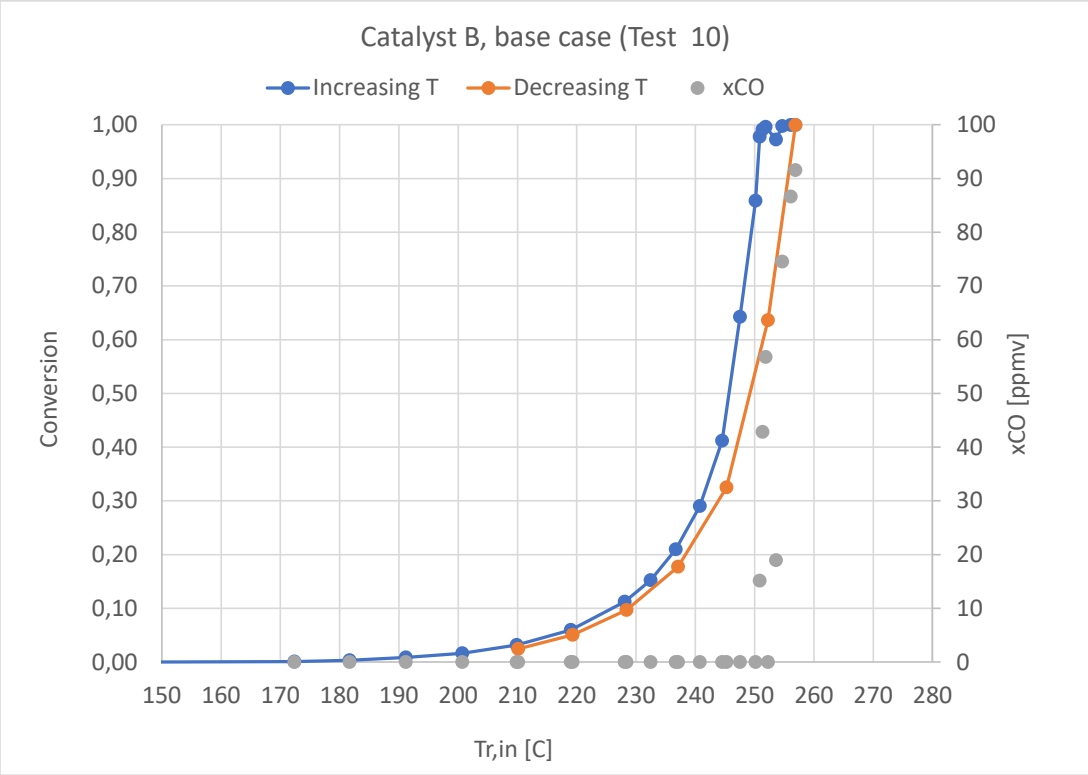


Findings catalyst A

- Full conversion (< 10 ppm O_2 at outlet) at ≈240 °C
- Wide hysteresis
- CO formation was only observed at temperatures higher than needed for full conversion

WP2 Oxygen Removal

Results catalyst testing

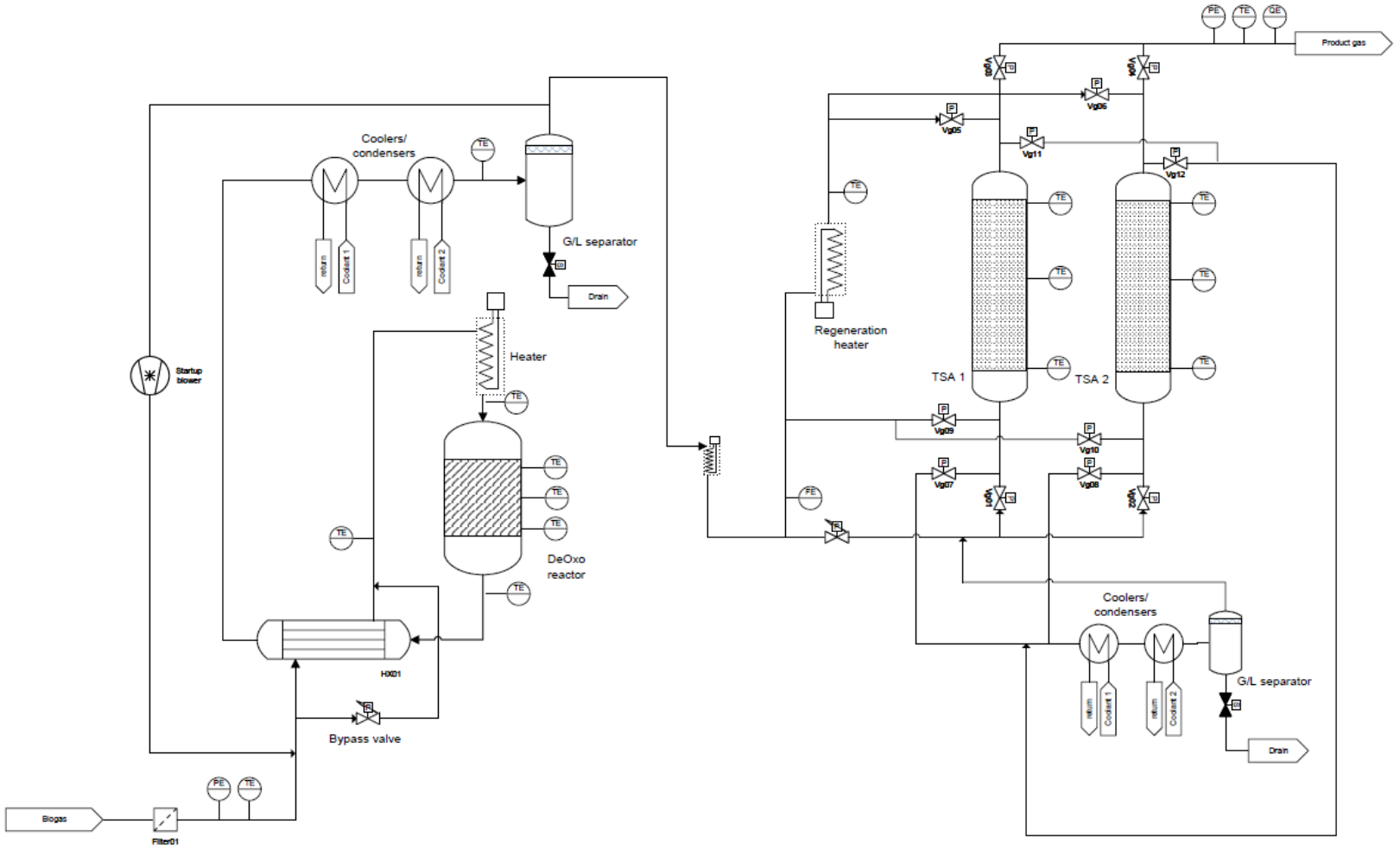


Findings catalyst B

- Full conversion (< 10 ppm O_2 at outlet) at higher temperatures than catalyst A
- No beneficial hysteresis and unstable oxygen conversion
- Considerable CO formation even when oxygen conversion is not complete

WP2 Oxygen Removal

Process design



WP2 Oxygen Removal

Conclusions & recommendations

Conclusions

- Preliminary process design focused on minimizing operational costs.
- < 10 ppm O_2 can be reached.
- Catalyst A is preferred over catalyst B due to lower temperature operation, wide hysteresis and minimal CO formation.

- Preliminary CAPEX estimate: €935,000.-
- Preliminary system footprint estimate: 4 x 3 m

Recommendations for future research

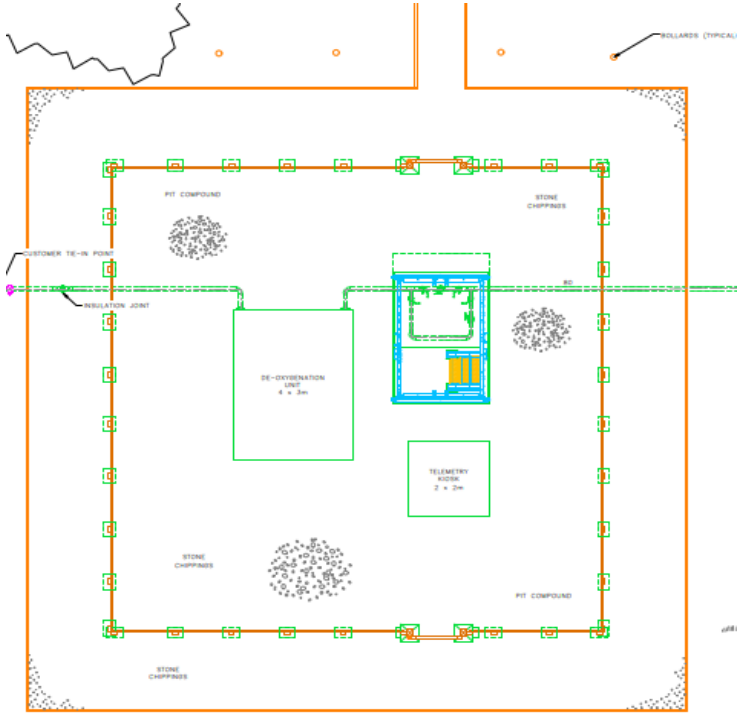
- Building a pilot unit for larger scale tests with catalyst A → this will also enable testing at higher pressure.
- Detailed process engineering to determine CAPEX and OPEX values.
- Business case evaluation (based on CAPEX and OPEX).

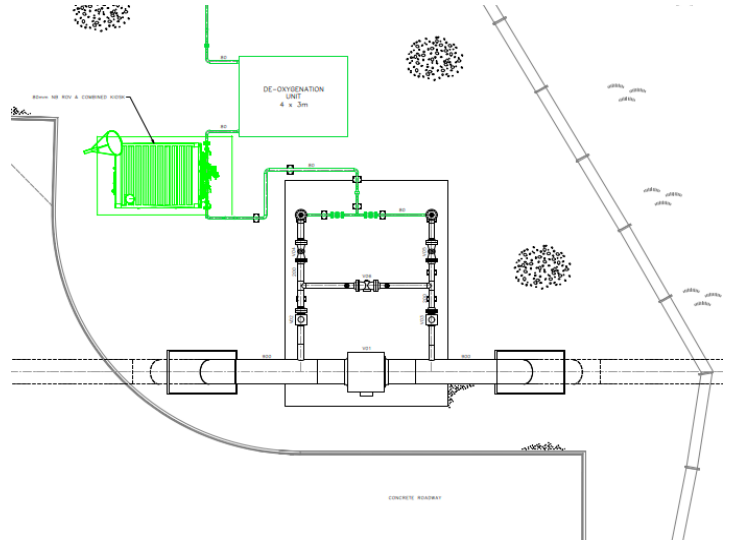
WP3 Storage Integration

- Change in philosophy from integration into storage facilities to biomethane entry points

- Two possible locations for integration of oxygen removal technology:
  - Biomethane Production Facilities
  - NGT Connection AGI

- There are challenges around locating the oxygen removal technology at either site due to:
  - Socialisation of costs
  - Contractual arrangements
  - Impact on design configurations





WP3 Storage Integration

Storage Site	Owner / Operator	Type	Working Gas Volume (declared)	Max. Withdrawal Rate (declared)	Max. Injection Rate (declared)
Aldborough	SSE Hornsea / Equinor	9 Salt Caverns	282 mcm	26 mcm/d	26 mcm/d
Hatfield Moor	Scottish Power	Depleted Gas Field	70 mcm	2 mcm/d	2 mcm/d
Hill Top	Kistos Energy Storage	5 Salt Caverns	59 mcm	13 mcm/d	13 mcm/d
Hole House Farm	Kistos Energy Storage	4 Salt Caverns (suspended)	-	-	-
Holford	Uniper UK	8 Salt Caverns	240 mcm	22 mcm/d	26 mcm/d
Hornsea (Atwick)	SSE Hornsea	8 Salt Caverns	308 mcm	12 mcm/d	3 mcm/d
Humbly Grove	Humbly Grove Energy	Depleted Oil Field	254 mcm	7 mcm/d	8 mcm/d
Rough	Centrica Energy Storage+	Depleted Gas Field	1,500 mcm	11 mcm/d	9 mcm/d
Stublach	Storengy UK	20 Salt Caverns	400 mcm	30 mcm/d	30 mcm/d

- Range of facilities in GB with the majority purpose build salt cavity based
- All are relatively long-established, with Stublach the most recent to come online in 2014.

- UESO provided engineering data for four sites: **Stublach, Aldborough, Atwick and Humbly Grove**
- Critical to oxygen removal considerations are injection rates, pressures and O_2 concentrations.
- Pipeline lengths exposed to 'wet gas' on studied sites ranged from **1km to 20km**
- Operational pressures in storage varied from **15 barg to 286 barg.**
- Pressures at NTS typically operating between **50-65 barg (for O_2 removal).**

- Challenges for O_2 removal at Storage Connection points:
  - **High Pressure**
  - **High Flowrate**
  - **Lower Concentrations** (compared to biomethane entry point)

- Challenges for corrosion inhibition at storage sites:
  - Range of pressure (including frequent cycling)
  - Length of affected piping systems
  - Chemical interaction with hydrate prevention dosing

- Focus on **removing oxygen at biomethane entry** points

WP4 Corrosion Prevention: Corrosion Threat Review

- Oxygen corrosion in wet conditions a **threat, even at low concentrations** - Levels **above 10 ppm is likely to elevate the threat.**
- Threat increases as **temperature reduces** near surface and **condensation** occurs.
- Example Corrosion Rate (96barg, 20 degC):
  - 0.2mol% O_2: **0.3mm per year**
  - 1 mol% O_2: **>1.5mm per year**
- **Oxygen levels < 10ppm** likely to reduce corrosion rate to **below 0.05mm per year.**
- Storage site pipework is designed with a **3mm corrosion allowance**. Potentially reducing lifespan to **2-10 years...**

- Most susceptible areas:
  - **Upper parts of Vertical subsurface pipelines** into wellheads
  - Pipelines between wellheads and dehydration facilities (buried & above ground) **close to wells**
  - **Water drop out at low points** in pipelines between wellheads and dehydration facilities

WP4 Corrosion Prevention : Mitigation Measures

Corrosion Inhibitors

- Limited options - **no 'tried and tested' solution found in the UK market** to reduce oxygen related corrosion in gas transmission pipelines
- Suppliers typically proposed alternative solutions (dehydration & oxygen removal)
- **US 'ChampionX' oxygen corrosion inhibitor available** on market but does not have regulatory approval for use in the UK - **Further R&D required**.

In-situ Internal Lining

- **'Batch' pigging applied internal lining** is possible but likely to be associated with significant challenges and uncertainties.
- Feasibility and cost evaluated on a **case-by-case basis**.
- High mobilization & set-up cost – Indicative estimate: **~£1.2M - 4.2M for 3-5km at diameters 300-750mm,** costs reduces thereafter.
- Spray applied solutions possible for shorter pipe sections only.
- **In-situ lining of vertical subsurface pipe** into storage caverns **unfeasible**.

WP4 Corrosion Prevention: Threat and Risk Management

Recommendations:

- **Site-specific corrosion assessment** considering specific pipeline design, historic and future operating envelope & **actual and expected Oxygen Content** (if not in place).
- Ensure **suitable corrosion mitigation is installed to monitor or mitigate** the threat. Ensuring that the oxygen content is below 10 ppm as a minimum.
- Review and revision of site **Risk Assessments**, especially **areas of increased susceptibility**. Updates to the Safety Case and Written Scheme of examination
- Develop a **long-term internal corrosion management plan** to monitor the threat (if not in place)

WP4 Corrosion Prevention: Suggested Further Study

Recommendations:

- **Further study of mitigation measures,** if deemed required by CBA and overall study outcome. This could include further investigation of Corrosion Inhibitors, understanding current status of testing and gaps to implement in the UK.
- Site specific **Contingency Planning** – e.g. Investigate the corrosion threat during an operational outage where the selected mitigation has failed. This would identify where corrosion would occur and quantify severity over a specific time period.
- **Consider performing a system wide risk screening across all storage sites:**
  - **Identify priority sites –** establish highest risk of future corrosion / highest corrosion rates to date.
  - Baseline the condition at highest risk sites through pigging & alternative inspection techniques – **confirm corrosion occurred to date.**
  - **Benefits:** Provide **reassurance** on the other storage sites & **demonstrate regulatory compliance.**

Risks

Ref	Risk Description	Likelihood (Low/Medium/High)	Impact (Low/Medium/High)	Mitigation
R10	Ownership of oxygen removal technologies and who will pay for them	High	High	Engage UESO early; discuss with ofgem on socialising of costs for oxygen removal technologies to enable more biomethane onto the NTS
R11	Oxygen removal flow rate is not suitable for storage sites at 2-30mcm/day	High	High	Consider deploying at biomethane entry points only.

Barriers

- 1% O2 HSE exemption for gas transmission objectives contradict with Project Remo2val objectives
- Conflict of interest with pit design for standardised biomethane connections project as oxygen removal deployment would require site to be an AGI.

Issues Resolved

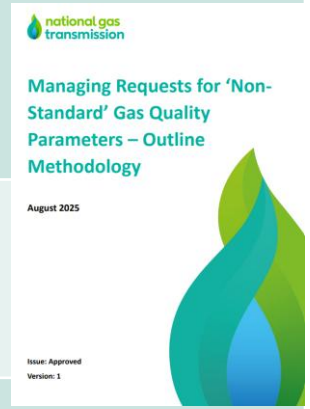
- Completing the project within 3-month timeframe and completing the testing of catalysts.

Update: Gas Quality – Non-standard requests

Transparent and Consistent: New Methodology

New methodology has been developed. This will be published on our website Connections Document Library

- Clear and transparent process documenting how we will manage requests for non-standard gas quality arrangements – oxygen (O_2) and Carbon Dioxide (CO_2)
- Non-Standard is that outside of our template Network Entry Agreement specification
- Methodology shaped through industry consultation via UNC Gas Quality Workgroup



- **Heat Map** of the NTS has been included in the methodology
  - Showing areas of the NTS where elevated O_2 may be accommodated more easily
  - Is based on likelihood of our pipeline conveying gas to "sensitive customers" within area
- **Network Analysis** will be undertaken for each application on a worst-case basis to determine risk
- Pre-defined **Decision-Tree**
  - Sets out clear steps to give a 'yes', 'no' or what level (of O_2/CO_2) can be accommodated
  - Methodology will be applied consistently
- Assessment made at the **Application** Stage (applicable to new connections)

For further information, or to share your thoughts on this subject, please contact Ahmed.jama@nationalgas.com

Qs still to Answer?

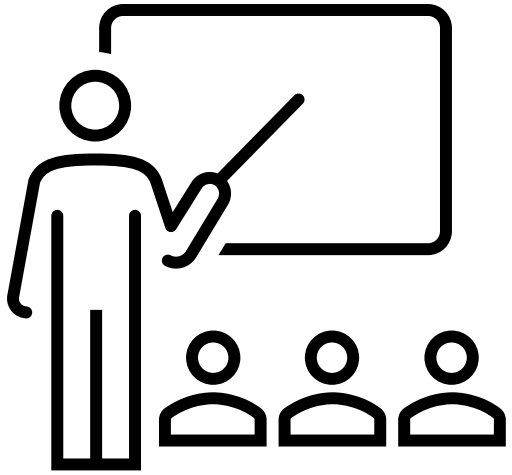
We will have Alignment and plan going forward with 1% oxygen HSE exemption if accepted or if modified to 0.5% or 0.6%. Current GSMR limit for O_2 at transmission level is 0.2%. However, Storage sites require 10ppm O_2 levels to prevent corrosion of their pipelines for wet gas environments.

Qs still to address:

- Is case by case network analysis for sensitive customers (e.g. power stations, storage sites, interconnectors) enough to show O_2 'blending out' for biomethane in the gas stream with decreasing natural gas demand and increasing biomethane on the NTS?
- Is there a point(at a %biomethane e.g. 50%) that the corrosion risk at storage sites becomes too great to not have Oxygen removal at every entry point?
- Do we need to start to deploy oxygen removal now and include in cost estimates for biomethane entry connections?
- Or can we wait until we reach a certain %biomethane in gas stream before we need to consider including oxygen removal?
- Also, only at new biomethane entry points or all existing biomethane entry points?'

Lessons Learnt

- Structuring of project around catalyst testing may not have been best approach and may have been better to have done the **economic modelling first** for business case. We have not looked into **electricity consumption** to gain accurate OPEX for oxygen removal.
- **Sharepoint access** from National Gas in a timely manner and organising that for all partners prior to Alpha
- Better **cross-fertilisation and workshops** throughout the project avoiding siloed work.
- Early **scope change not fit for purpose** and definition of change with early identification that oxygen removal must be deployed at biomethane entry points.
- Better **scope definition and communication** around 3-month timeframe
- **Flexibility** with innovation project scopes to alter as project progresses and new findings come to light
- Took time to get **early RFIs**, accelerate engagement early on.
- Start with a **scoping workshop**, this is what we need and initial scope of project for comments and input.



Project Specific Conditions

Condition 1 The Funding Party must not spend any SIF Funding until contracts are signed with the Project Partners named in Table 1 for the purpose of completing the Project.

Condition 2 The Funding Party must report on the financial contributions made to the Project as set out in its Application. Any financial contributions made over and above that stated in its Application should also be reported and included on the Innovation Funding Service (IFS).

Condition 3 The Funding Party must make reasonable endeavours to participate in all meetings related to the Project that they are invited to by Ofgem, UKRI and the Department for Energy Security and Net Zero during the Discovery Phase.

Condition 4 The Discovery phase will last for a period of up to five months from the date the Project Direction is issued; the Project will be allowed a flexible start date within the five-month period. The Project must provide the monitoring officer with the start date of the Project and must be completed before the end of the five-month period.

Comms and Engagement Plans

- **Show and Tell** to be scheduled for early March
- **Linkedin Posts** – NGT, HyGear, Premtech and FI plan to do project completion posts. UESO will publish on their website a post.
- Internal NGT update to SMEs as part of **Innovation Asset Working Group** (IAWG)
- **Biomethane forum** engagement with biomethane producers

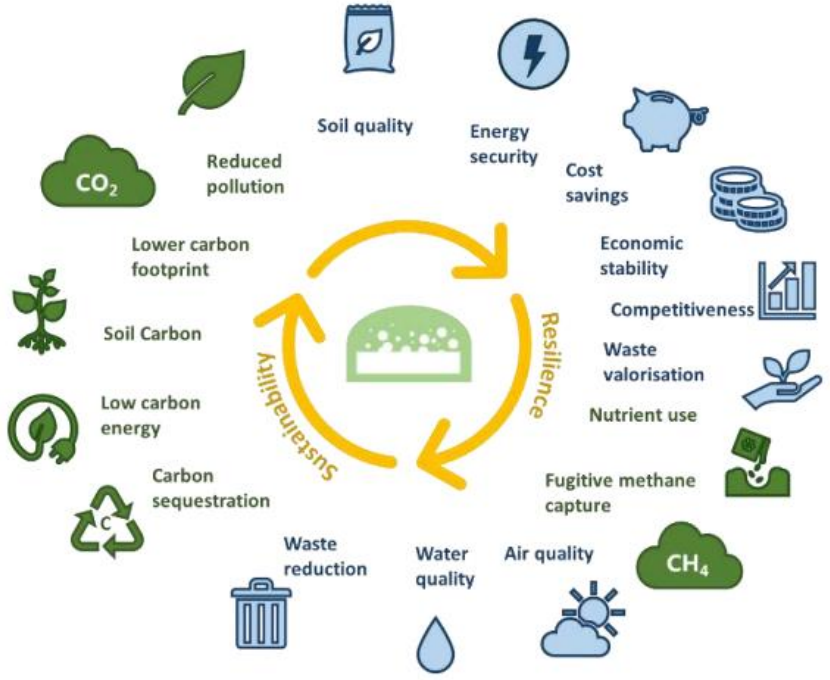
Alpha and plans for what's next

We have decided to **delay any application to Alpha until cycle 6** or beyond to give us sufficient time to complete the NIA **standardised biomethane connections** project and have a **HSE decision on the 1% oxygen exemption** for gas transmission hopefully in **March** which this project outcomes impact

Qs still to answer?

Who pays/funds for the 10ppm oxygen removal technology?

Do we only need to deploy oxygen removal near sensitive customers or at every entry point? How do we align with 1% oxygen exemption submitted to HSE and not contradict aims of methodology presented to them.



Thank you

national gas transmission

Project RemO2val

SIF Discovery Phase

Design Report

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Issue	Description	Originator	Date	Checker	Date	Approver	Date
01	SIF Discovery Phase Issue	N. Horan-Martin	09/01/26	L Frearson	12/01/26	N. Horan-Martin	15/01/26
00	Issue for internal review	N. Horan-Martin	15/12/25	L Frearson	08/01/26		

Revision Index

Revision	Summary of Changes
00	Issue for Internal Review
01	Updated with internal and client comments

Abbreviations

CDM	Construction (Design) and Management
DSEAR	Dangerous Substances and Explosive Atmosphere Regulations
GEU	Grid Entry Unit
GS(M)R	Gas Safety (Management) Regulations
HAZID	Hazard Identification
HAZOP	Hazards in Operability
IGEM	Institute of Gas Engineers and Managers
LOPA	Layer of Protection Assessment
MEC	Minimum Entry Connection
NB	Nominal Bore
NCC	National Control Centre
NIA	Network Innovation Allowance
NTS	National Transmission System
ppmv	Parts Per Million by Volume
PSR	Pipeline Safety Regulations
PSSR	Pressure System Safety Regulations
SIF	Strategic Innovation Fund
SIL	Safety Integrity Level
SWDS	Safe Working Desing Study
UK	United Kingdom

Executive Summary

The use of greener gases, such as biomethane, are an important part of the UK's transition to net zero. Storage sites are critical for balancing seasonal supply and demands for energy. However, increased levels of oxygen (up to 0.2% molar) in biomethane can lead to corrosion of storage facility assets in wet gas conditions, compromising their integrity. This project will conduct a comparative analysis to evaluate the technical and economic viability of advanced catalytic and adsorption technologies for reducing oxygen levels to 10 ppm or lower in biomethane. The study will also assess the use of corrosion inhibitors to ensure the integrity and longevity of critical storage infrastructure.

Premtech's role within the project is to integrate the advanced catalytic and adsorption oxygen removal technology, developed by project partner HyGear, into existing biomethane infrastructure. This includes design considerations for two integration scenario's depending on the final asset ownership of the oxygen removal technology. The scenarios are as follows:

- Integration within the 3rd party owned biomethane production facility.
- Integration within the National Gas owned biomethane entry connection.

The final decision regarding ownership of the oxygen removal technology will be made by National Gas Transmission, based on stakeholder engagement (considering aspects such as socialising costs across a number of parties) and a cost benefit analysis.

Contents

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1 Project Background

The use of greener gases, such as biomethane, are an important part of the UK's transition to net zero. Storage sites are critical for balancing seasonal supply and demands for energy. However, increased levels of oxygen in biomethane can lead to corrosion of storage facility assets in wet gas conditions, compromising their integrity. This project will conduct a comparative analysis to evaluate the technical and economic viability of advanced catalytic and adsorption technologies for reducing oxygen levels in biomethane. The study will also assess the use of corrosion inhibitors to ensure the integrity and longevity of critical storage infrastructure.

1.1 Project Scope

The project will:

- **Implement advanced catalytic oxidation and adsorption technologies** to reduce oxygen levels to below 10 ppm, meeting industry safety standards.
- **Integrate oxygen removal units** with existing biomethane infrastructure, ensuring seamless operation and continuous monitoring.
- **Or apply corrosion inhibitors and protective coatings** to safeguard storage facilities.
- **Conduct regular inspections and maintenance** to proactively manage and mitigate corrosion risks.

The RemO_2val project directly supports the Strategic Innovation Fund (SIF) Innovation Challenge on green gases by addressing a critical barrier to the safe and scalable deployment of biomethane: oxygen-induced corrosion in underground storage systems. RemO_2val facilitates the safe and efficient storage of biomethane, a key green gas, by developing and deploying advanced oxygen removal technologies. This ensures that biomethane can be reliably stored and transported through existing infrastructure, accelerating the UK's transition to a net-zero energy system.

By reducing oxygen to levels below 10 ppmv, the project significantly mitigates the risk of corrosion in wet gas environments, thereby extending the lifespan and reliability of underground storage assets. This aligns with SIF's goal of ensuring a resilient and future-ready gas network.

The project's focus on preventing corrosion translates into substantial cost savings by reducing the need for maintenance, repairs, and asset replacement. These savings support the delivery of low-carbon energy solutions at the lowest possible cost to consumers, aligning with a core SIF objective.

RemO_2val's modular design and integration strategy allow for scalable deployment across the UK's gas infrastructure. The project also incorporates real-time monitoring and control systems, ensuring compliance with safety standards and enabling safe operation at scale.

1.2 Premtech Scope

Premtech's role within the project is to integrate the advanced catalytic and adsorption oxygen removal technology, developed by project partner HyGear, into existing biomethane infrastructure. This includes design considerations for two integration scenario's depending on the final asset ownership of the oxygen removal technology. The scenarios are as follows:

- Integration within the 3rd party owned biomethane production facility.
- Integration within the National Gas owned biomethane entry connection.

The final decision regarding ownership of the oxygen removal technology will be made by National Gas Transmission based on stakeholder engagement (considering aspects such as socialising costs across a number of parties) and a cost benefit analysis.

1.3 Biomethane Customer Connections

1.3.1 Business As Usual

Typically, biomethane connections to the National Transmission System (NTS) are completed using a Minimum Entry Connection (MEC). The MEC is generally comprised of an isolation valve (remotely or locally operated depending on risk assessment), full bore bypass with isolation valve, pressure monitoring and vent / purge points. Figure 1 and Figure 2 below present typical remote and locally operated biomethane connections.

It should be noted that any Remotely Operated Valve (ROV) is not a safety device and is generally only closed at the request of the customer.

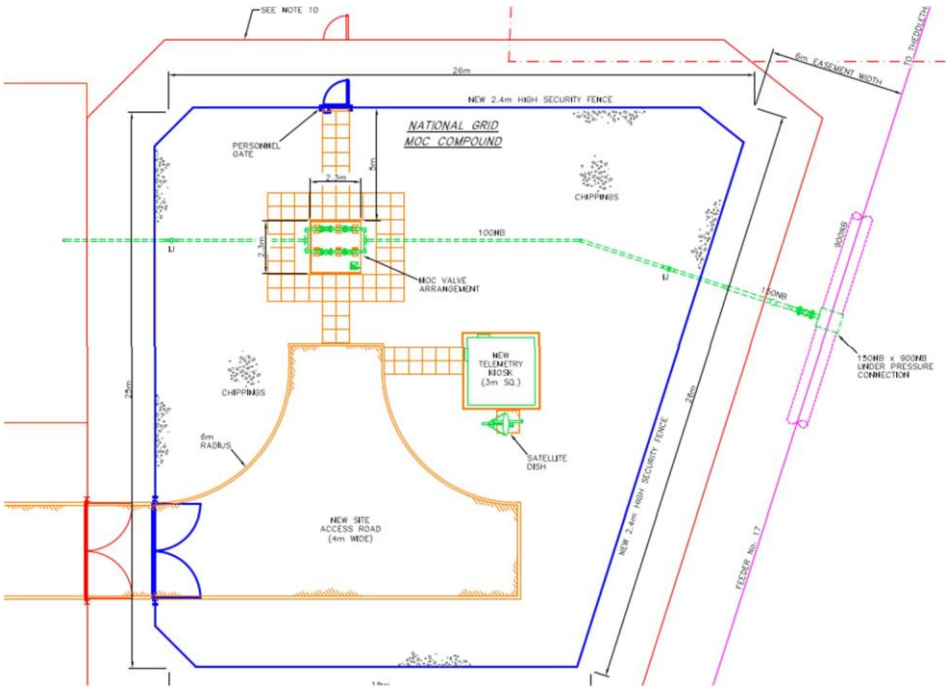

Figure 1 - MEC Connection with ROV

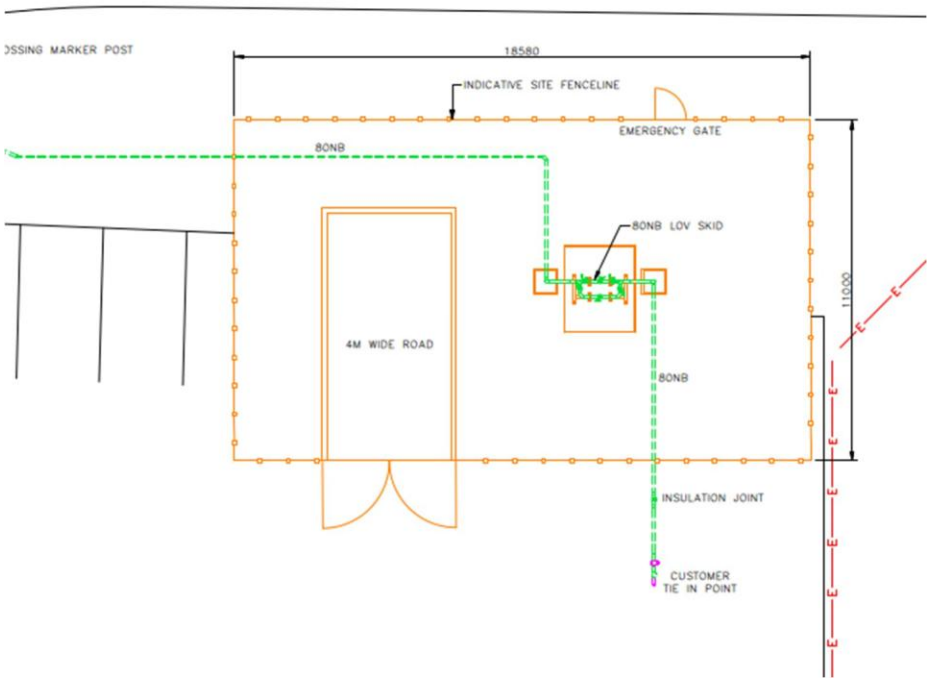

Figure 2 - MEC Connection with LOV

1.3.2 Innovation Project

National Gas are currently developing a 'standard MEC in a pit' arrangement for biomethane entry connections using Network Innovation Allowance (NIA) funding. This design is intended to become the standard MEC offering for future biomethane connections.

The provides biomethane producers with a more cost-effective connection option compared to current MEC designs, as the equipment housed within a covered pit. This arrangement reduces the need for a security fence, site road and other ancillary infrastructure, thereby reducing overall land take. Figure 3 below illustrates the design developed to date

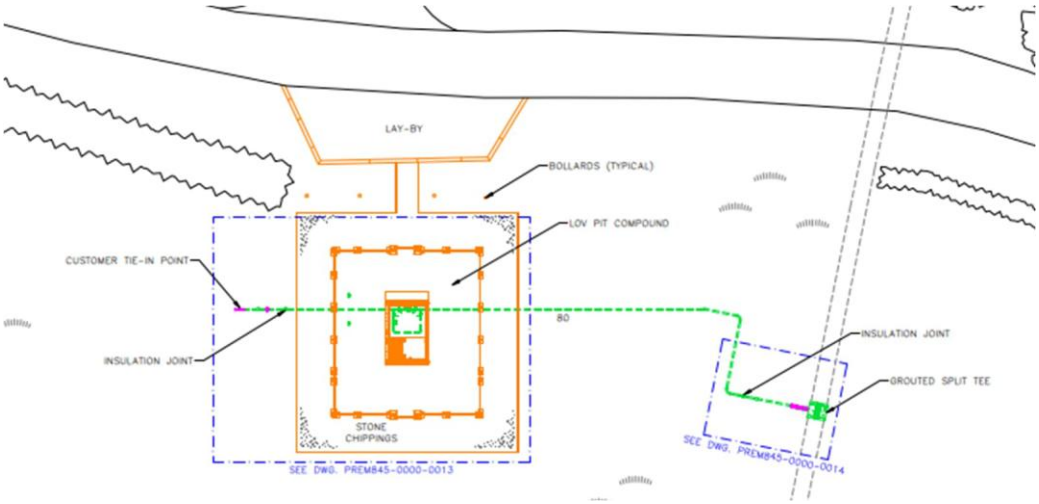

Figure 3 - Standard MEC in a pit arrangement

2 Design Requirements

2.1 Legislation

For the design of the works, full consideration has and will be given to the following and other relevant legislations, where applicable. These include:

- The Construction (Design and Management) (CDM) Regulations 2015 [1]
- Health and Safety at Work etc. Act 1974 [2]
- Pipeline Safety Regulations (PSR) 1996 [3]
- The Pressure Systems Safety Regulations (PSSR) 2000 [4]
- Dangerous Substances and Explosive Atmospheres Regulations (DSEAR) 2002 [5]
- Gas Safety (Management) Regulations (GS(M)R) 1996 (with 2023 amendments) [6]

2.2 Specifications

Where appropriate, the work elements are principally designed in accordance with the following specifications and standards from the Institution of Gas Engineers and Managers (IGEM) and National Gas Transmission.

- IGEM/TD/13 Edition 3 [7] – Pressure Regulating Pipeline Installations Exceeding 7 bar
- IGEM/SR/25 Edition 2 [8] – Hazardous Area Classification of Natural Gas Installations
- T/SP/PW/11 Part 1 [9] – Pipework Systems Operating at Pressures Exceeding 7 bar Part 1 – Design and Materials
- T/SP/TR/18 [10] – Engineering of Pipelines and Installations Operating at Above 7 bar

2.3 Gas Safety (Management) Regulations

Schedule 3 of GS(M)R [6] sets out the maximum allowable limits for various contents that may be contained within the natural gas transported on the NTS. The maximum allowable content of oxygen, as defined in GS(M)R [6], is 0.2% (on a molar basis) when the operating pressure of the pipeline is greater than 38 barg.

3 Design Considerations

3.1 Oxygen Removal Technology

The oxygen removal technology being developed by HyGear is expected to have a footprint of approximately 4m x 3m (refer to email dated 08/12/25 reproduced in Appendix A). However, this is still subject to the final unit design, which will be completed in subsequent project phases (subject to successful funding applications).

Based on preliminary line sizing calculation completed by HyGear, it has been agreed that the connecting flange size will be 40NB with a class 600 pressure rating (suitable for ~100 barg), please refer to email dated 06/11/2025 in Appendix A.

3.2 Technology Location

3.2.1 Biomethane Producers

Biomethane production sites include the anaerobic digester, which produces the biomethane, and the necessary post-processing plant (clean up, compression etc.) to ensure the produced biomethane is within the limits imposed by GS(M)R 1996 [6] for components such as oxygen and hydrogen sulphide. Therefore, there is a logicical case for the producers to own and operate the oxygen removal technology.

The biomethane production facilities typically consist of:

- Packaged solution for gas cleaning / upgrading,
- Grid Entry Unit's (GEU's),
- compression, and
- metering.

The GEU confirms the biomethane is within specification and can be injected into the NTS; therefore, the oxygen removal technology would need to be installed prior to the GEU and its associated gas quality system.

The exact location of the oxygen removal technology would need to be confirmed on a site-by-site basis depending on the post-processing required and the equipment / packages needed to meet the post-processing requirements, recycle / reject line locations and gas quality measurement points. PRZ0028-0000-ELD-0001 reproduced in Appendix A has indicatively shown the oxygen removal technology between the upgrade plant and GEU for a provided scheme.

3.2.2 National Gas Connection

As the current oxygen limit within GS(M)R 1996 [6] is 0.2%, and the desired oxygen level to avoid corrosion in wet gas conditions at storage sites is lower at 10 ppmv, then there may be resistance to installing the oxygen removal technology on biomethane producers' sites. Producers could reject this requirement as it exceeds existing contractual arrangements and will likely increase their operational costs.

In this case, there would be a need to install the oxygen removal technology within the National Gas owned compound. However, this approach introduces several challenges.

By needing to install the oxygen removal technology within the National Gas compound, the ability to deploy the 'standard MEC in a pit' arrangement would be diminished. As the oxygen removal technology would need to be installed above ground, a security fence would be required to prevent third-party interference with any above ground assets.

Furthermore, there would also be a need to install gas analysis equipment to verify that the oxygen content is within the allowable limit of 10ppmv. Both the oxygen removal technology and gas analysis equipment would require power and have signals (such as oxygen content) to transmit via a telemetry unit housed in a telemetry kiosk.

PRZ0028-GEN-1000-0010 reproduced in Appendix B shows the potential layout of the oxygen removal technology including the 'standard MEC in pit'. However, given the need for a telemetry kiosk, power supply, security fence etc., it is considered unlikely that a pit arrangement would be acceptable because of the increased health and safety risks of operating within a pit. This defeats the purpose of the 'standard MEC in a pit' arrangement whereby planning permission is achieved quicker along with reduced planning costs and overall site footprint. Therefore, the connection would need to be at an AGI.

The current, typical MEC with ROV (at an AGI) already includes for the provision of a telemetry kiosk, power supply and security fence making it the more practical arrangement for deploying the oxygen removal technology. A potential layout for this configuration is shown on PRZ0028-GEN-2000-0010 reproduced in Appendix B.

Therefore, it is suggested that the need for oxygen removal is included as a decision for consideration within the decision tree for the 'standard MEC in a pit' arrangement and clarity is required if oxygen removal is placed on every entry point or only near sensitive customers, i.e., storage facilities.

If oxygen removal units only need to be deployed near sensitive customers, then this would enable the 'standard MEC in a pit' arrangement to still be utilised more widely, especially for smaller producers to help encourage biomethane growth by reducing their costs and enabling planning permission being granted quicker.

3.3 Remote Monitoring

Continuous monitoring of oxygen content in the supplied biomethane is essential to ensure compliance with the 10 ppmv tolerance limit. This will be achieved via an on-site telemetry unit.

If the oxygen removal technology is incorporated within the biomethane producers' compound, the telemetry unit would also be located there, with signals repeated via 4G (subject to a coverage survey) to a 'donor' telemetry unit at an existing National Gas owned AGI. This reflects the typical arrangement currently utilised for biomethane connections.

The telemetry unit will receive the necessary gas quality signals from either the oxygen removal technology and / or the gas quality analysis equipment and will then repeat the signals

to the National Gas' National Control Centre (NCC). The signals will be transmitted to NCC circa every four (4) minutes, which is the typical sample time of a gas chromatograph.

High and High-High alarm limits are likely to be set within the telemetry unit to alert NCC if the oxygen content approaches the allowable limit of 10 ppmv. The high high alarm may also trigger an 'executive action' such as the closure of a valve to prevent the injection of biomethane until the oxygen content returns to an acceptable value. alternatively, there may also be the opportunity to maintain the supply of biomethane into the NTS and blend out the higher oxygen content subject to factors such as proximity to a storage site, volumes of gas and oxygen contents on those gas streams.

An initial, indicative cause and effect matrix has been produced and is reproduced in Appendix B.

4 Health and Safety

4.1 General

Health and safety issues will be addressed throughout the design and construction of the project. The modifications will be constructed in accordance with current health and safety legislations, including the Health and Safety at Work Act, the Construction (Design and Management) Regulations 2015 (CDM) and the management of Health and Safety at Work Regulations.

4.2 Hazardous Areas

Hazardous area drawings have been produced for both the 'standard MEC in a pit' arrangement and the current, typical MEC arrangement. It should be noted that as the oxygen is removed via catalytic or adsorption means there are no hazardous areas generated for oxygen and the zones generated are solely for natural gas.

There are no continuous or primary sources of grades of release. All hazardous areas generated are secondary grades of release from flanges, screwed fittings, joints, valve glands and regulators.

As the Maximum Operating Pressure (MOP) of the connection is 94 barg and conditions are normal, the zones applied are taken from Table 1 of IGEM/SR/25 [8] as follows:

Table 1 - Hazardous Area Zonings

Operating Pressure (bar)	Zoning Distance (m)
94	1.5

4.3 Formal Process Safety Assessments

There will be a need to conduct suitable FPSA's for the inclusion of the oxygen removal technology. The scope of the required studies will be confirmed by NGT in a T/PM/HAZ/9 proforma and will likely include the following:

- Hazard Identification (HAZID)
- Hazards in Operability (HAZOP)
- Safe Working Desing Study (SWDS)

As the oxygen removal technology is reducing the oxygen content from GS(M)R compliant levels to a preferred concentration for storage sites, it is unlikely that the oxygen removal units will need a Safety Integrity Level (SIL) or Layer of Protection Assessment (LOPA), but this will be confirmed by the HAZOP.

5 Bibliography

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[2] HSWA, *Health and Safety at Work etc. Act 1974,* legislation.gov.uk, 1974.

[3] PSR, *Pipeline Safety Regulations (PSR) 1996,* legislation.gov.uk, 1996.

[4] PSSR 2000, *The Pressure System Safety Regulations 2000,* legislation.gov.uk, 2000.

[5] DSEAR 2002, *Dangerous Substances and Explosive Atmospheres Regulations (DSEAR) 2002,* legislation.gov.uk, 2002.

[6] GS(M)R, *Gas Safety (Management) Regulations 1996 with 2023 amendments,* HSE, 2023.

[7] IGEM/TD/13 Edition 3, *Pressure Regulating Installations for Transmission and Distribution,* IGEM, 2023.

[8] IGEM/SR/25 Edition 2, *Hazardous Area Classification of Natural Gas Installations,* IGEM, 2010.

[9] T/SP/PW/11 Part 1 , *Specification for Pipework Systems Operating at Pressures Exceeding 7 bar Part 1 - Design & Materials,* National Gas, 2024.

[10] T/SP/TR/18, *Specification for Engineering of Pipelines and Installations Operating at Above 7 barg,* National Grid, 2021.

Appendix B Premtech Documentation

Document Title	Document Number	Included (Y/N)
Cause and Effect Matrix	PRZ0028-SCH-0000-0001	Y

Project RemO2val

Document Number - PRZ0028-SCH-0000-0001 Revision A (05/01/2025)

Premzero

Biomethane Entry Site			Effect	Description	Alarm to GNCC	Valve Close (Note 2)	Alarm to GNCC	Alarm to GNCC	Alarm to GNCC	Alarm to GNCC	Alarm to GNCC	Alarm to GNCC	Alarm to GNCC
Cause and Effect Matrix				GNCC Tag	TBC	Valve 001	Valve 001	TBC	Mains1	Mains2	RTU Fault	Intruder 1	Charger 1
Cause				Action	On	On	On	On	On	On	On	On	On
Rev	Description	Site Tag	Logic										
A	High Oxygen Content	TBC	TBC	1	X								
A	High High Oxygen Content	TBC	TBC	1	X	X (Note 1)							
A	Valve fails to move to correct position	Valve 001	1oo1	1			X (Note 1)						
A	Gas Analysis Fault	TBC	TBC	1				X					
A	Phase Failure	N/A	1oo1	1					X				
A	AC Power Supply Failure	N/A	1oo1	1						X			
A	RTU Fault	N/A	1oo1	1							X		
A	Intruder Detection	N/A	1oo1	1								X	
A	Battery Charger Failure	N/A	1oo1	1									X

Note 1: Only applicable to Remotely Operated Valves

Note 2: Requirement for valve closure t

GENERAL REPORT

Jorn Heinst

HYG-PR000020-205-RP.001

HYGEAR

Table of Content

1 Introduction

National Gas Transmission operates and maintains high-pressure pipelines that transport natural gas from production and import points to distribution networks and major industrial users, ensuring safe and reliable delivery across the country.

National Gas Transmission is investigating the technical and economic viability of using catalysts and corrosion inhibitors to mitigate the effects of corrosion on the storage infrastructure. Please find a brief project summary below:

The use of greener gases such as biomethane are an important part of the UK's transition to net zero. Underground storage sites for biomethane are critical for balancing seasonal supply and demands for energy. However, increased levels of oxygen in biomethane can lead to corrosion of assets in wet gas conditions, compromising the integrity of storage facilities. This project will assess in a comparative analysis the technical and economic viability of advanced catalytic and adsorption technologies to reduce oxygen levels in biomethane with corrosion inhibitors to ensure the integrity and longevity of critical storage infrastructure.

National Gas Transmission has engaged HyGear for the deoxygenation and drying part of this project (catalytic and adsorption technologies).

2 Scope

The purpose of the project is to remove the oxygen from biomethane so that it can be stored in underground storage sites without corrosion compromising the integrity of the storage facilities. HyGear will determine the most effective way to do this by selecting the proper deoxygenation catalyst and doing some preliminary process engineering.

HyGear deliverables

1. Report detailing the testing procedure (set-up, process conditions, etc.) and reporting the test results. The tests in this first project will not be performed in the 40 – 70 bar(g) pressure range, but the results will be extrapolated to predict the performance within this specified pressure ranged based on previous higher-pressure tests and theoretical modelling based on literature.
2. A preliminary process flow diagram detailing our process from the inlet (gas from the NTS) to the outlet (dried gas with oxygen content < 10 ppmv, suitable for underground storage) of our system.
3. Report describing the total project approach and the assumptions made. The process design will also be described (to provide some further explanation to the process flow diagram). This report will also contain HyGear's recommendation for the catalyst selection considering: a) the experimental results, b) the total process design and c) economic considerations.

It is important to note that the placement of the HyGear DeOxo system has changed since these deliverables were drafted to make the flow rate a better fit for the HyGear technology. The system will be placed downstream of a biomethane producer's upgrading system and upstream of the injection point into the National Gas grid. For this reason, the design data (pressure, O_2 content, etc.) have also changed. Please find the relevant design data in the next chapter.

This report is deliverable 3 and will also discuss findings from the test report (deliverable 1) and will describe the PFD in more detail (deliverable 2).

This project is primarily research-focused, as the available budget does not allow for detailed process engineering. Activities typically associated with such engineering—such as the development of detailed P&IDs, heat and mass balances, equipment sizing, and system layouts—are therefore outside the project scope. Consequently, the test report constitutes the key deliverable. Any process engineering presented in this report is preliminary in nature and is limited to the development of a process flow diagram (PFD), based on current knowledge and the outcomes of the catalyst testing.

3 Design data

The original design data is listed below.

Original design specifications

- The feed gas to our system has specifications in line with the NTS (National Transmission Grid), which means that the pressure will be in the 40 – 70 bar(g) range with a dew point of < -10 °C.
- The oxygen content can be as high as 1 vol% (10,000 ppmv) O_2.
- The oxygen content needs to be reduced to below the < 10 ppmv limit required by the UESO (Underground Gas Storage Operators) and the gas needs to be dried.

However, the relevant, up-to-date design data for the biomethane producer case is listed below. This data will be used from now on. These are the average values. The requirement of < 10 ppmv O_2 at the outlet of the deoxygenation and drying system is unchanged.

Table 1: Average process conditions and composition

Specifications of biomethane production at Somerset farm		
Flow rate	785	Nm^3/hr
Pressure	59	bar(g)
Temperature	17	°C
Composition		
O_2	0.1	vol%
H_2O (dew point)	-43.8	°C
H_2S	2.2	ppmv

A more detailed composition can be found in Annex 1. Please keep in mind that the values in Annex 1 come from one sample, while the values above are averages.

The H_2S content of 2.2 ppmv is expected to be too high for the selected catalysts, as the supplier recommends to stay below 0.5 ppmv. During a potential follow-up study, solutions will be proposed.

Even though the figures above have been kept in mind during the preliminary process design and the experiments, they will not return in this report as detailed process engineering is not part of the project scope (see the explanation in the previous chapter).

4 Process design and catalyst selection

This chapter presents the overall process design of the deoxygenation and drying system, followed by a recommendation for the selected catalyst. It concludes with a summary of the economic considerations. As noted earlier, detailed process engineering is outside the scope of this project. Accordingly, the process description provided in this chapter is preliminary and qualitative in nature, and is based on HyGear's current knowledge and the results of the (current and previous) experimental work. As a result, the economic analysis is also qualitative in nature.

4.1 Process description

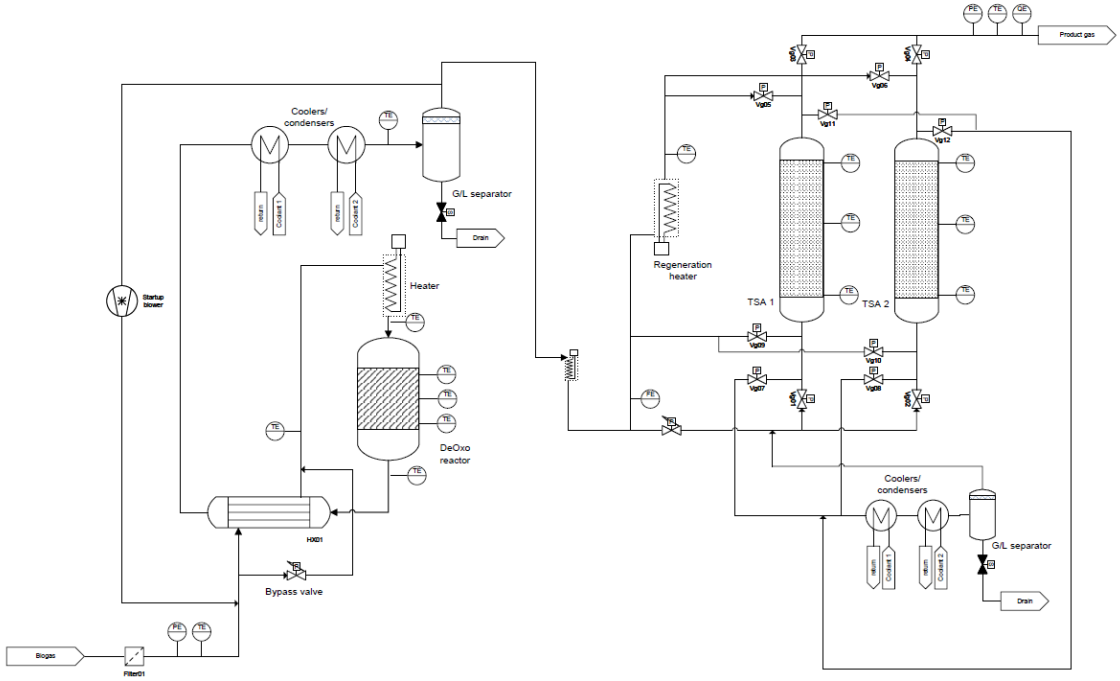


Figure 1: Preliminary PFD of the deoxygenation and drying system

The incoming gas is first passed through a strainer to remove any coarse particles. It is then preheated in a recuperative heat exchanger (HX01), using the hot product gas from the DeOxo reactor. Subsequently, the gas is heated to the required DeOxo reactor inlet temperature by an electrical heater, which is integrated into the DeOxo reactor.

The DeOxo reactor is filled with a noble-metal catalyst. In the reactor, oxygen reacts with methane, resulting in an exothermic reaction and a corresponding temperature increase. The progress of the reaction is monitored using temperature sensors installed along the catalyst bed. The reactor outlet gas is cooled in several steps: first in HX01 against the incoming feed gas, followed by two sequential coolers to condense as much water as possible. This two-step cooling strategy minimizes energy consumption: the first cooler is operated with coolant from a dry cooler, while the second cooler uses coolant from a chiller. After condensate separation, the gas is slightly reheated to prevent downstream condensation and is then sent to the TSA (Temperature Swing Adsorption) system for further drying. A start-up blower is included in the DeOxo system to enable reactor heat-up by recycling and heating gas during start-up.

The TSA system removes residual water vapor by adsorption on a zeolite. It consists of two parallel vessels filled with adsorbent. At any given time, one vessel is in adsorption mode,

drying the feed gas, while the other is either being regenerated or is in standby. By switching the appropriate valves, the feed and regeneration flows are routed to the correct vessel for each phase of operation.

During adsorption, the gas flows from bottom to top through the vessel. For vessel 1, valves Vg01 and Vg03 are opened; for vessel 2, valves Vg02 and Vg04 are opened. Regeneration is carried out by diverting a portion of the biomethane stream, heating it in the regeneration heater, and passing the hot gas from top to bottom through the adsorbent bed. The regeneration flow is routed via valves Vg05 and Vg07 for vessel 1, and Vg06 and Vg08 for vessel 2. The hot gas leaving the vessel bottom is cooled in two steps, and the resulting condensate is removed in a gas–liquid separator. The dry gas is recycled back into the main feed line upstream of the TSA, ensuring that no valuable methane is lost.

After the hot regeneration phase, the bed is cooled using a portion of the feed gas. In this cooling phase, the gas flows from bottom to top through the vessel, controlled by valves Vg09 and Vg11 for vessel 1, and Vg10 and Vg12 for vessel 2. The hot outlet gas is again routed through the coolers and condensate separator, similar to the regeneration phase. The dew point, temperature, and pressure of the dried product gas are continuously monitored.

4.2 Catalyst recommendation

As described in the test report, two noble-metal catalysts (catalyst A and catalyst B) were evaluated experimentally. Based on the results, HyGear recommends selecting catalyst A for use in the deoxygenation and drying process. The key considerations supporting this recommendation are summarized below. Please refer to the test report for a more detailed analysis.

Catalyst A achieves full oxygen conversion at a lower temperature than catalyst B: 100% conversion is reached at approximately 240 °C, compared to 250–260 °C for catalyst B. Operating at a lower temperature offers two important advantages:

- Lower energy consumption: Heating and cooling of the biomethane stream are among the largest contributors to the overall electricity demand. Consequently, a system using catalyst A is expected to have lower electrical power consumption.

- Extended catalyst lifetime: Catalyst deactivation over time leads to an increase in the temperature required to maintain the desired conversion. Starting from a lower operating temperature provides a greater margin, resulting in a longer catalyst lifetime and fewer costly catalyst replacements.

Alternatively, instead of operating at a lower temperature, it could also be decided to operate at a higher temperature but use less catalyst. This has a positive impact on the CAPEX, as the catalyst is one of the major cost drivers.

In addition, catalyst A exhibits a pronounced hysteresis between increasing and decreasing temperature, which is not observed for catalyst B. During temperature decrease, catalyst A maintains near-complete conversion down to significantly lower temperatures. This behaviour enhances operational flexibility, as short-term temperature drops caused by process disturbances do not immediately result in a loss of conversion. However, it should be noted that the long-term behaviour of the catalyst is still unknown, so this advantage might become less pronounced over time.

Finally, CO formation is a critical consideration, as CO is a strong catalyst poison that can deactivate the catalyst and CO is also an unwanted byproduct. For catalyst A, CO formation only occurs at temperatures well above those required for full oxygen conversion. With appropriate heat management, catalyst A can therefore be operated at full conversion for extended periods without producing CO. In contrast, catalyst B shows CO formation at significantly lower temperatures, already occurring at or near the temperature required for full oxygen conversion, making it less suitable for long-term stable operation. Moreover, the conversion of oxygen is unstable, probably necessitating even higher operational temperatures accompanied by still higher CO levels.

4.3 Economic considerations

The preliminary process design of the deoxygenation and drying system has been developed with a strong focus on minimizing both capital expenditure (CAPEX) and operational expenditure (OPEX), while ensuring reliable long-term operation and product gas quality.

From an operational cost perspective, energy consumption and catalyst replacement are the dominant contributors. The largest energy demands arise from heating the biomethane to the DeOxo reactor operating temperature and from cooling and condensation of water downstream of the reactor. The use of a recuperative heat exchanger significantly reduces the required electrical heating duty by recovering heat from the reactor outlet gas. In addition, the application of two-stage cooling—using a dry cooler followed by a chiller—minimizes the use of chilled utilities and thus reduces electricity consumption. The TSA system is designed to regenerate the adsorbent using a recycled fraction of biomethane, ensuring that no valuable methane is vented or flared, which further reduces operating costs. Considering the foreseen operation at high pressure, the amount of water in the gas after condensation will be low, consequently the TSA will not consume a lot of energy.

Catalyst selection has a direct impact on both energy consumption and maintenance costs. As discussed in Section 4.2, catalyst A achieves full oxygen conversion at lower temperatures than catalyst B. This reduces the required heating and cooling duty of the system, leading to lower electricity consumption over the lifetime of the plant. Furthermore, operating at a lower temperature provides additional margin against catalyst deactivation, potentially extending catalyst lifetime and reducing the frequency of catalyst replacement. Since noble-metal catalysts represent a significant portion of the CAPEX and replacement costs can be substantial, this has a meaningful impact on lifecycle costs.

Another important economic factor is process reliability and availability. The favourable hysteresis behaviour of catalyst A allows the system to tolerate temporary temperature disturbances without an immediate loss of conversion, reducing the risk of off-spec gas production and unplanned shutdowns. Moreover, the delayed onset of CO formation for catalyst A lowers the risk of catalyst poisoning, contributing to longer uninterrupted operating periods and lower maintenance costs.

While the inclusion of multiple heat exchangers, coolers, and a dual-vessel TSA system increases the initial equipment count, this is offset by reduced energy usage, minimal methane losses, and improved operational robustness. Overall, the selected process configuration and the use of catalyst A is expected to result in the lowest total cost of ownership when considering CAPEX, OPEX, and long-term operational reliability, based on our experience and the catalyst

testing results. However, as already mentioned at the start of this chapter, this can not yet be quantified as a more detailed process engineering study would be required.

5 Summary

The deoxygenation and drying system is designed to efficiently remove oxygen and water from a biomethane stream while minimizing operational costs (energy consumption, frequency of catalyst replacement and methane losses). The process combines heat integration, staged cooling, and a zero-loss temperature swing adsorption (TSA) system to achieve high product quality with robust and reliable operation. Heat from the DeOxo reactor outlet is recovered to preheat the incoming gas, and staged cooling allows water to be condensed efficiently with minimal use of chilled utilities. The TSA system ensures further drying while recycling regeneration gas, preventing the loss of valuable biomethane.

Experimental evaluation of two noble-metal catalysts identified catalyst A as the preferred option. Catalyst A achieves complete oxygen conversion at lower temperatures than catalyst B, which reduces electricity demand for heating and cooling and potentially provides a greater margin against catalyst degradation. Its favourable hysteresis behaviour allows the system to tolerate temporary temperature fluctuations without immediate loss of conversion. Moreover, CO formation—which acts as a catalyst poison—occurs only at temperatures well above normal operating conditions, enabling long-term operation without catalyst poisoning. Catalyst B, by contrast, forms CO at or near the temperatures required for full oxygen conversion, increasing the risk of catalyst deactivation and operational interruptions.

From an economic perspective, energy consumption and catalyst replacement are the primary operational cost drivers. Operating at lower temperatures with catalyst A reduces both electricity use and maintenance requirements, while the rest of the process was designed to minimize methane losses and electricity consumption and ensure high availability. The inclusion of multiple heat exchangers, coolers, and TSA vessels increases initial equipment investment but is offset by lower operational costs and improved long-term reliability. Overall, the system configuration with catalyst A provides an energy-efficient, cost-effective, and operationally robust solution for biomethane deoxygenation and drying, offering the lowest total cost of ownership and the most reliable long-term performance.

Key Takeaways

- Catalyst selection: Catalyst A achieves full oxygen conversion at lower temperatures and minimizes the risk of CO poisoning.
- Energy efficiency: Heat recovery, staged cooling, and TSA regeneration with recycled gas reduce electricity consumption and methane losses.
- Operational reliability: Catalyst hysteresis and robust process design allow stable operation under temporary disturbances, reducing downtime and off-spec production.
- Economic benefits: Lower energy use, extended catalyst life, and minimal methane losses reduce OPEX, while the system design ensures high availability, resulting in the lowest total cost of ownership.

6 Future research recommendations

As the DeOxo reactor, and thus the catalyst, is the core of the system, future development work should focus on this. As stated in the test report, HyGear recommends to pursue further development work based on the application of catalyst A. The test results indicate that a reaction design based on a GHSV in the range 5000-10000 h^{-1} should be feasible; this determines the required catalyst bed volume for a given feed flow rate. HyGear also recommends to perform catalyst tests at higher pressure.

As indicated earlier, detailed process engineering has not been part of this engineering study, and the process design developed to date is qualitative in nature. HyGear therefore recommends a follow-up study (Alpha phase), in which a quantitative process design can be developed based on process data provided by the biomethane producer. This would also include the addition of a H_2S removal step. This would result in a detailed process design—including P&IDs, heat and mass balances, and system layouts—as well as a comprehensive cost assessment covering both CAPEX and OPEX. When these CAPEX and OPEX figures are known, the business case can be evaluated to determine whether oxygen removal at the biomethane producer's site would be economically feasible.

During the project, the following (non-binding) estimates have been shared by HyGear:

- Turnkey system cost: €935,000.-
- System footprint: 4 x 3 m

These estimates would be turned into firm, binding figures during the follow-up study.

Annex I Somerset Farm AD Plant sample

	Units	Values
Date sampled		29/06/2020
Time sampled		12:50
Composition:		
H_2S	ppmv	< 1
CO_2	mol%	0.07
O_2	mol%	0.14
N_2	mol%	3.48
H_2	mol%	< 0.01
CH_4	mol%	96.30
C_2H_6	ppmv	< 1
C_2H_4	ppmv	< 1
C_3H_8	ppmv	< 1
C_4H_{10}	ppmv	< 1
C_5H_{12}	ppmv	< 1
C6 molecules	ppmv	< 1
C7 molecules	ppmv	< 1
C8 molecules	ppmv	< 1
C9 molecules	ppmv	< 1
C10 molecules	ppmv	< 1
C11 molecules	ppmv	< 1
C12 molecules	ppmv	< 1
Hydrocarbon Dew Point	°C	-135


DEOXO LAB TEST REPORT

Jan Peter Brouwer

HYG-PR000020-205-RP.002

HYGEAR



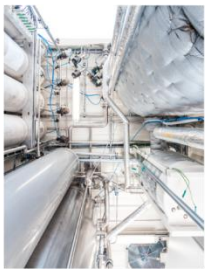
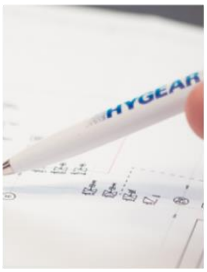



Table of Content

1 Introduction

This report describes the results of catalyst tests, that have been performed in HyGear's laboratory as a part of the pre-engineering study for National Gas into the deoxidation of biomethane. Typical analyses of a biomethane stream at two different moments, as provided by National Gas are given in Table 1. The oxygen content is in both cases around 0.15%, but in other cases it may be up to 1%, therefore a concentration of 1% is taken as the maximum O_2 concentration in this work.

Table 1: Typical biomethane compositions (source: e-mail K. Shillinglaw dd 09-10-2025)

Parameter	**Units**	**Value** **sample gas 1**	**Value** **sample gas 2**
H_2S	ppmv	<1	<1
CO_2	mol%	0.070	1.37
O_2	ppmv	1400	1500
N_2	mol%	3.48	0.91
H_2	mol%	<0.01	<0.01
CH_4	mol%	96.3	97.56
C_2H_6	ppmv	<1	<1

Deoxidation of biomethane occurs by the reaction between oxygen and methane over a catalyst according to the following reaction equation:

$$CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O$$

Typically, supported noble metal catalysts are used to carry out this reaction, which usually occurs at temperatures above 200°C. During the reaction heat is released, which causes an increase of the temperature: in pure methane, an oxygen content of 1% leads to a temperature increase of 83°C if heat losses are absent. The temperature increase is approximately proportional to the feed oxygen concentration. It is also pointed out that the gas composition changes somewhat due to the reaction: if 1% of O_2 is present in the feed gas then at complete conversion 1% of reaction water will be added to the product gas (on top of the water present in the feed). This may be significantly more than the water originally present in the feed (which could be of the order of 0.1-0.2%, but likely less). Moreover, up to 0.5% of CO_2 will be added by the reaction to the product gas. This may be comparable to the CO_2 content of the feed gas (although this appears to fluctuate somewhat).

In order to meet specifications for storage and transportation, the treated biogas should have an oxygen content < 10 ppmv, whereas the feed content could be as high as 1% (=10.000 ppmv) as mentioned above. This means that high conversions of up to 99.90% are required. Aim of the laboratory tests is to understand the relation between the obtained conversion and the process conditions (temperature, gas composition, space velocity). The test data may serve as a basis for reactor and system design. It is noted that due to limitations of the test setup, the tests have been performed only at pressures close to atmospheric, whereas in the field pressures up to 70 bar may be encountered.

Two commercial noble metal catalysts have been tested, denoted here A and B. These contained different noble metals on a support, at a content of 0.25-0.30 wt%. HyGear has previously performed laboratory deoxidation tests with these catalysts with varying simulated

biogas compositions. Results of those tests are proprietary and confidential. These results have been used for designing a number of deoxidation systems for flow rates up to 2500 Nm^3/h.

The current tests are performed with pure CH_4 from gas cylinders. Oxygen is supplied by means of compressed air; this adds some nitrogen to the gas mixture as well, which presumably behaves inert. According to the gas analyses from Table 1 the N_2 content in biomethane may be in the range 0.5-4%, which corresponds quite well with the values obtained by the air addition in the tests. In addition, the CO_2 content of biomethane may be up to 1.5% but it may also be much lower; in the current tests no CO_2 is added to the feed gas. Furthermore, the dew point is specified as <-10°C. The water content then depends on the total pressure, and at atmospheric pressure this dew point corresponds to a maximum water content of 0.25%; at higher pressures the water content will be lower. In the tests the maximum water content in the feed will be 0.25%; the water formed by the reaction will usually be more than this. Water might have a somewhat slowing effect on the reaction, therefore the maximum content is used in the tests in order to remain conservative.

2 Test setup

2.1 Test reactor

The tests are performed with a small reactor that contains approximately 45 ml of catalyst (corresponding to a catalyst mass of about 34 g). A sketch and a photograph of the test reactor are shown in Figure 1.

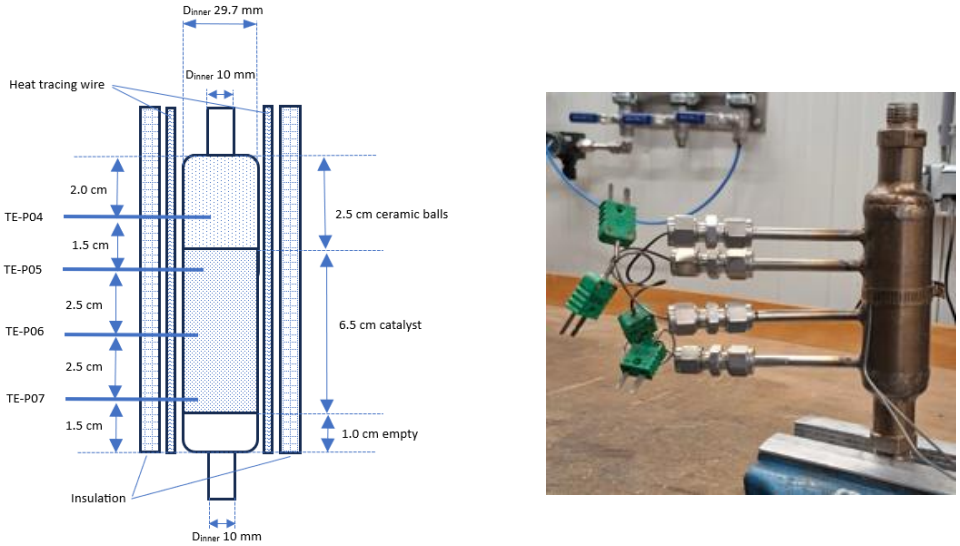


Figure 1: Skecth and photograph of the test reactor (flow direction is from top to bottom)

During a test, the flow direction is from top to bottom. On top of the catalyst a layer of ceramic balls is present that serves to distribute the incoming flow; the ceramic balls rest directly on the catalyst (no separation gauze). The ceramic balls have a diameter of 3 mm, whereas the catalyst has a size distribution of 2-4 mm (spheres). The bed rests on a support grid that is fixed in the lower part of the reactor. The reactor is externally insulated, and contains heat tracing wire underneath the insulation for compensating heat losses (the heating power of the tracing is controlled by means of a thermocouple at the external reactor wall). Thermocouples are placed through welded side-tubes at different positions inside the reactor, before and in the catalyst bed, to measure the temperature profile.

The total gas flow rate through the reactor in a test is typically either 4 or 8 slm, corresponding to a GHSV of 5290 and 10600 h^{-1} respectively. Superficial velocities at actual conditions are in the range 0.1-0.4 m/s.

2.2 P&ID of the setup

The P&ID is shown in Figure 2.

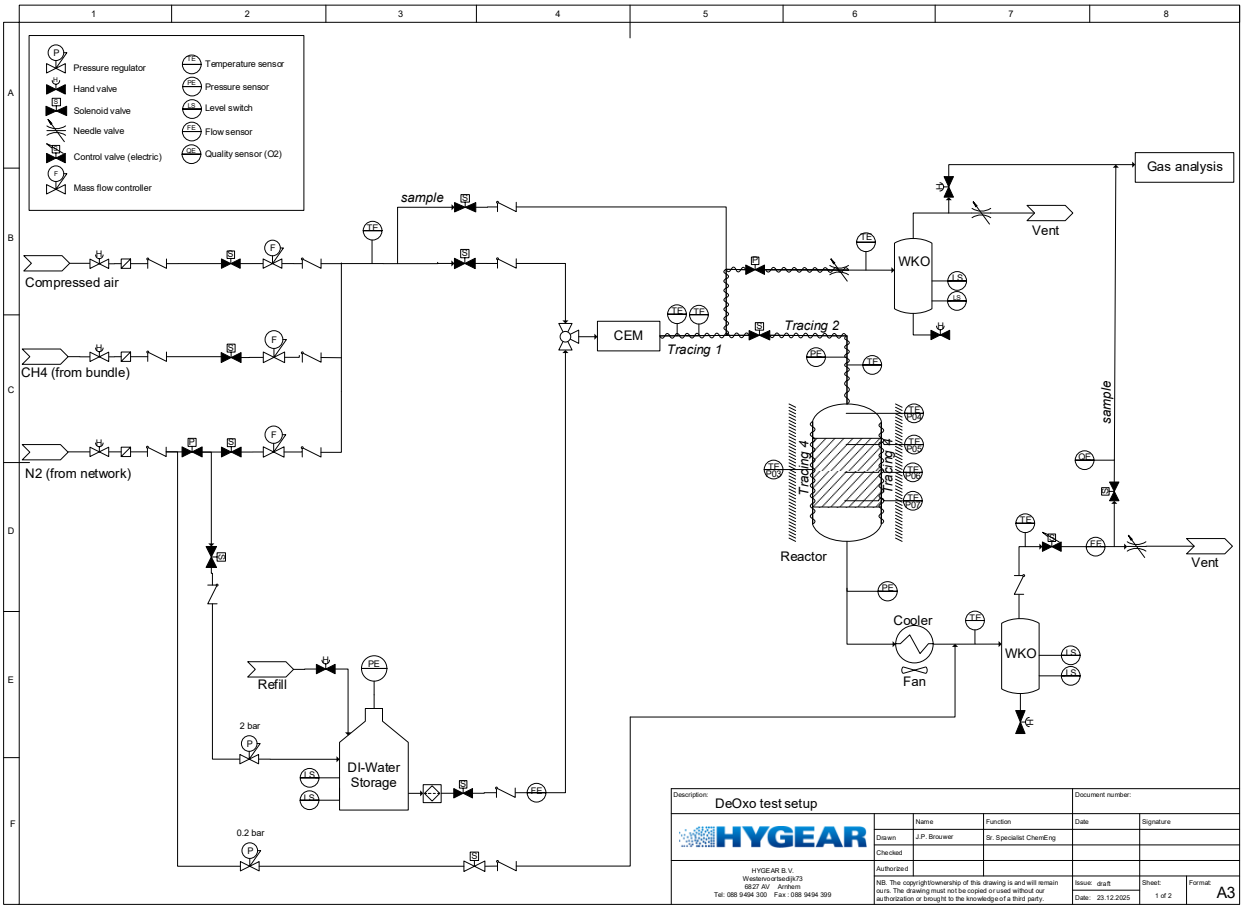


Figure 2: P&ID of the test setup (simplified)

The feed gas mixture consists mainly of (pure) methane from gas cylinders with oxygen added in the form of air from HyGear's compressed air network. The gases are supplied into the system by means of mass flow controllers. Optionally the gas can be humidified by means of a CEM-unit (=Controlled Evaporator Module); this module doses liquid water with a mass flow controller and manages its complete evaporation and mixing with the carrier gas. The liquid water is forced to the CEM by applying nitrogen pressure to the water storage tank. In principle all the methane+air is also forced through the CEM unit. The CEM-unit is heated internally to well above the dewpoint of the gas mixture. After the CEM-unit the gas is heated by means of several pieces of heat tracing wire to the desired inlet temperature, and then fed into the reactor. As noted before, the reactor itself is also heated by means of a separate heat tracing section. The gas from the reactor is cooled down by forced air cooling, and potentially formed condensate is separated in the WKO vessel (=water knock-out). A control valve downstream of the WKO allows to regulate the reactor pressure; the maximum pressure is limited to 2 bar(a). Most tests are performed at 1.2 bar(a), some at 1.8 bar(a).

After the WKO a sample flow is sent to the gas analyzers (O_2-sensor, CO_2/CO analyzer). The O_2-sensor (oxyIQ) can measure down to 1 ppmv, depending on the selected range.

The total gas flow can be in the range of 2 to 10 slm (in these tests it was typically either 4 or 8 slm).

Figure 3: The test setup in HyGear's laboratory

2.3 Test procedure

The test procedure is as follows: before each test the setup is first flushed with nitrogen for several minutes. Then the CH_4 and air MFC's are set to the selected values for the test, and the nitrogen is switched off. After a few more minutes the CEM unit is set as well (if applicable). Next, heating is started by means of the heat tracing wires, until the reactor wall temperature reaches 150°C.The reactor temperatures are recorded, as well as the O_2 concentration in the product gas (at this point, the latter is virtually equal to the feed concentration, since the temperatures are yet too low for the reaction to occur). When these are all constant, the tracing temperature setpoint is increased in steps, and after each step the settings are kept constant until the reactor temperatures and O_2 concentration are stable. When reaction occurs, CO_2 appears in the product gas, and the measured CO_2 concentration should be stable as well.

The temperature stepping is continued until the measured oxygen concentration in the product gas has turned to <10 ppmv; then usually one more temperature step is done. Depending on the catalyst activity the temperature steps can be smaller or larger; usually in the region where the reaction rates change fast, the temperature is increased in steps of 5°C.

Next the temperature is decreased again in steps. It is already mentioned here that one of the catalysts showed a considerable hysteresis, whereas the other one showed almost equal conversion curves for increasing and decreasing temperature.

As explained before, the reactor wall temperature is used as the setpoint for the heat tracing (TE-P03 in the P&ID). In discussing the results the conversions will be related to the reactor inlet temperature (TE-P04). In an adiabatic reactor this is an obvious choice, however the test reactor is not truly adiabatic since heat losses are relatively high though these are partly compensated by heat tracing.

3 Test results

3.1 Test programme

Once the reactor is charged with catalyst several tests are performed; in each test the conversion versus temperature is recorded for a fixed set of conditions, as explained in section 2.3. A full test typically takes 1-2 days. Following parameters are varied in the tests:

- **Flow rate**
  The total feed flow rate should be considered in relation to the weighed-in amount of catalyst; this is often expressed as the space velocity or GHSV (=Gas Hourly Space Velocity), which is defined as the flow rate at standard conditions and on an hourly basis divided by the catalyst bed volume. In general, at a higher flow rate (and so at a higher GHSV) a higher temperature will be needed to reach a certain conversion. In the tests, the total feed flow rate is either 4 or 8 slm.
- **Oxygen feed content**
  The maximum O_2 content in the application is specified as 1%. Tests are performed with two O_2 contents, i.e. 0.5% and 1.0%. Since the maximum product O_2 concentration is specified as 10 ppm (fixed), the required conversion is higher at higher feed concentration (99.90% vs. 99.80%). At equal total flow rates, more O_2 has to be converted if the O_2 concentration is higher, which may require higher temperatures; on the other hand, the temperature increase across the reactor will also be higher at higher feed concentrations.
- **Pressure**
  As pointed out, the pressure of the test setup is limited to a maximum of 2 bara. Most tests are performed at a pressure of 1.2 bara; a few tests are performed at an increased pressure of 1.8 bara. This is still very far from the maximum pressure of 70 bara that can be encountered in the actual application, nonetheless it represents a 50% increase in the absolute pressure and therefore also a 50% increase in residence time. The pressure also affects the adsorption of O_2 on the catalyst surface; at higher pressure the O_2 partial pressure increases and the surface coverage increases as well (but when the surface is already virtually fully covered then obviously there will be hardly an effect of increasing pressure).
- **Water content**
  The dew point of biomethane is <-10ºC. At atmospheric pressure this corresponds to a water content of 0.25%; this is the very maximum, and in the tests where water is added this maximum amount is taken. In a number of tests the water is left out.

The parameters were varied systematically with respect to the base case, which was defined as follows:

- Flowrate 4 slm
- O_2 content 1%
- Water content 0.25%
- Pressure 1.2 bara

A flow rate of 4 slm gives a quite conservative GHSV, while the O_2 content is maximum and the water content as well.

By varying one parameter at a time compared to the base case, the influence of each parameter can be assessed. On top of this a few extra combinations were tested as well (i.e. more than

one parameter was varied). Table 2 lists the conditions in all the tests; the base case test numbers are 4 and 10.

Table 2: Overview of test conditions

Test #	Catalyst	Flow rate [slm]	O2 content [mol%]	Pressure [bara]	H_2O content [mol%]	Variation wrt base case
1	A	4	0.5%	1.2	0%	$x_{O2}\downarrow$, $x_{H2O}\downarrow$
2	A	4	0.5%	1.2	0.25%	$x_{O2}\downarrow$
3	A	4	1.0%	1.2	0%	$x_{H2O}\downarrow$
4	A	4	1.0%	1.2	0.25%	-
5	A	4	1.0%	1.8	0.25%	P↑
6	A	8	1.0%	1.2	0.25%	Flow↑
7	A	8	1.0%	1.8	0.25%	Flow↑, P↑
8	B	4	0.5%	1.2	0.25%	$x_{O2}\downarrow$
9	B	4	1.0%	1.2	0%	$x_{H2O}\downarrow$
10	B	4	1.0%	1.2	0.25%	-
11	B	4	1.0%	1.8	0.25%	P↑
12	B	8	1.0%	1.2	0.25%	Flow↑
13	B	8	1.0%	1.8	0.25%	Flow↑, P↑

The weighed-in amount of catalyst was about 33.5 g for both catalyst, corresponding to a bed volume of approximately 45 ml. A total flow rate of 4 slm corresponds to a GHSV of 5340 h^{-1}, whereas a total flow rate of 8 slm this is 10700 h^{-1}.

3.2 Results of base case tests

Figures 4 and 5 show the conversion vs. temperature plots for the base case for each catalyst. As explained before, in the test the temperature is first increased in steps, and when the conversion is complete the temperature is decreased in steps; the curves for increasing and decreasing temperature are shown separately. In addition, the measured CO-concentrations are shown in the plots (grey dots, right axis).

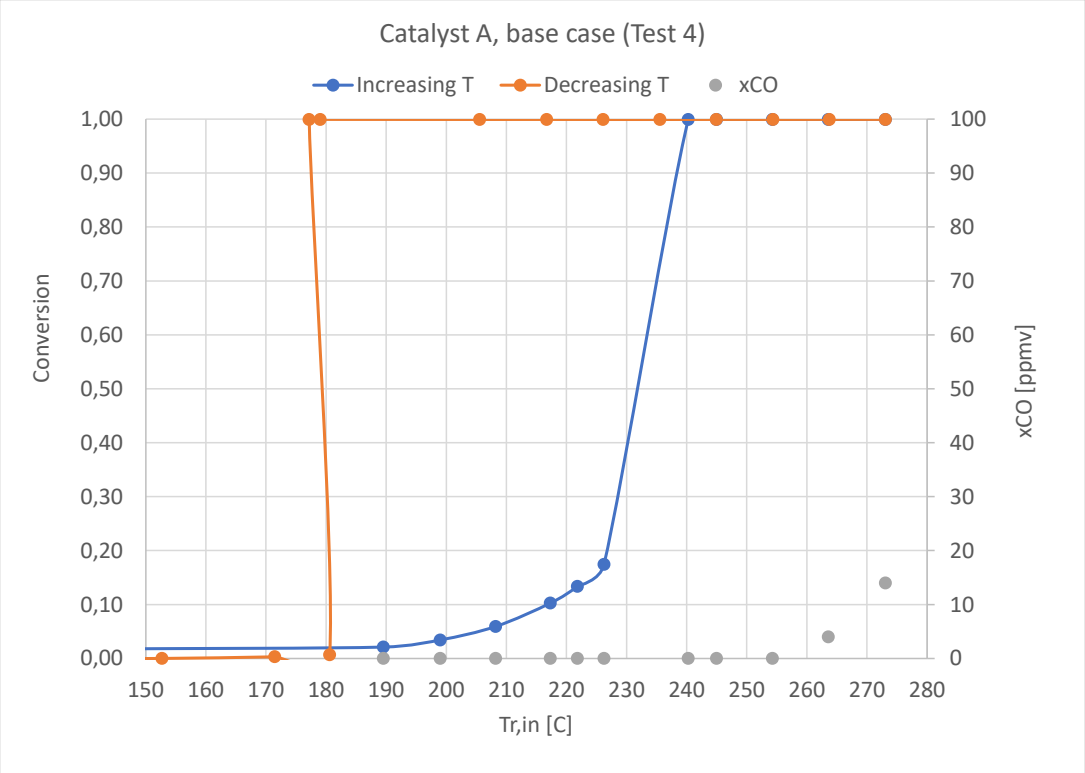


Figure 4: Results of base vase test for Catalyst A (flow 4 slm / 1.0% O_2 / 0.25% H_2O / 1.2 bara)

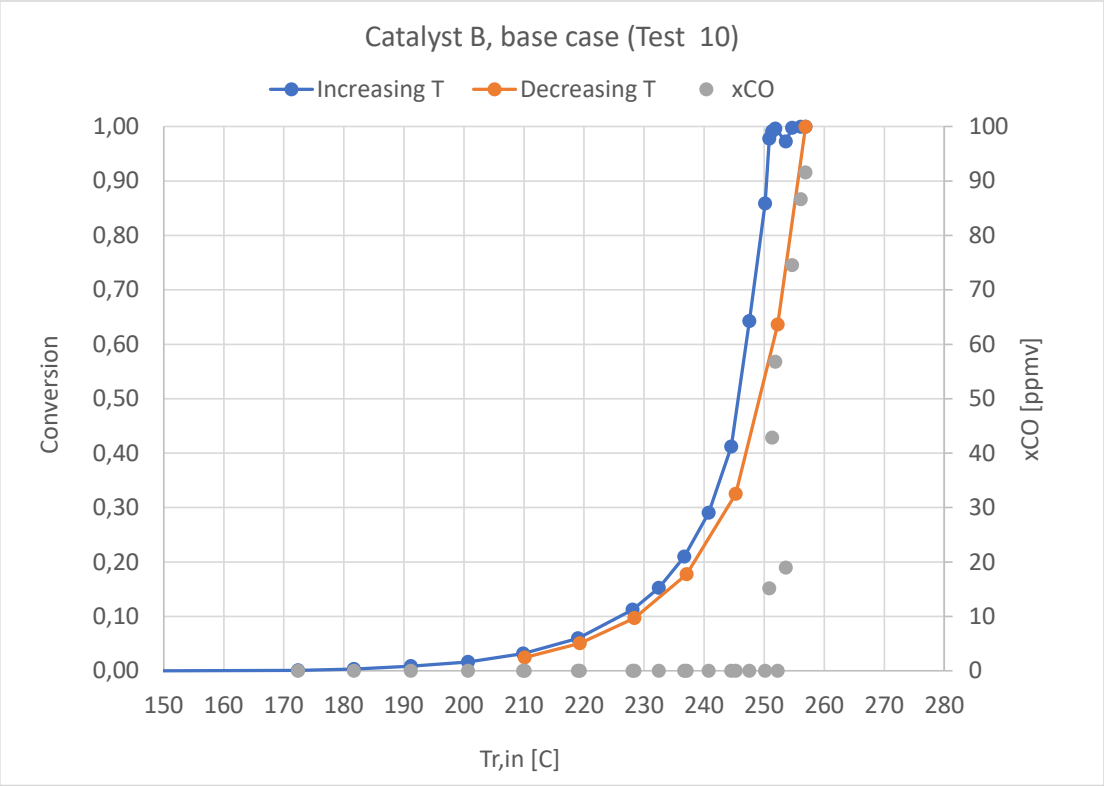


Figure 5: Results of base case test for Catalyst B (flow 4 slm / 1.0% O_2 / 0.25% H_2O / 1.2 bara)

In the graphs it can be seen that the conversion changes a lot in just a small temperature window. Catalyst A shows a wide hysteresis between increasing and decreasing temperature, which is not observed with catalyst B. For catalyst A the lowest temperature at which complete conversion is reached is about 240ºC for the increasing temperature curve vs. about 180ºC for decreasing temperature (note: at the points where the conversion is close to 1 the measured O_2 concentrations were below 10 ppmv). The testing time for each temperature step appeared to be an important factor. At the lowest temperature with full conversion on the decreasing temperature curve, the test conditions were kept stable for 4 h with no O_2 detected by the sensor; but during this period some temperature changes were noted in the bed, i.e. the temperature at the middle of the bed decreased (TE-P06) while the temperature at the end of the bed increased (TE-P07), pointing to widening of the reaction zone. Towards the end of the 4 h the bed temperatures looked stable. See figure 6, test period between the 1st and 2nd dashed lines. Upon subsequent further decrease of the reactor temperature the O_2 concentration in the product increased. It took up to 11 h before the O_2 concentration became stable and equal to the input concentration; the bed temperature at the end of the bed (TE-P07) decreased in steps, accompanied by stepwise increase in the O_2 concentration in the product gas. The final increase concerned a sudden big leap from about 2000 to 10000 ppm within a short time. During this leap the inlet temperature (TE-P04) increased a little, which was due to the automatic controls, that increased the tracing power due to the ceasing of the production of reaction heat. See figure 6, period between the 2nd and 3rd dashed lines. No CO formation was detected in this period.

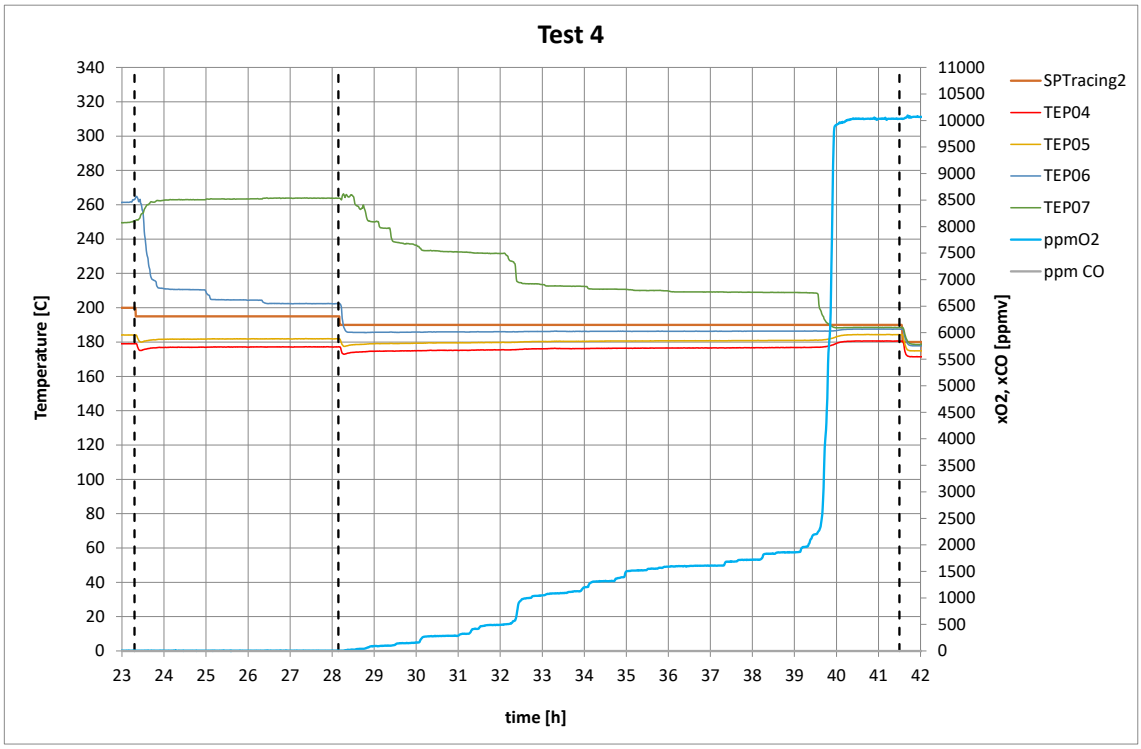


Figure 6: Temperature and concentration profile development during Test 4 (base case, Catalyst A).

With catalyst B it was experienced that at intermediate to high conversions (>50%), the O_2 concentration in the product was quite unstable. In between the dashed lines in figure 7 (covering a period of 10 h) the temperatures were constant, yet the O_2 content in the product

gas increased and fluctuated up and down. There appeared to be a correlation with the CO content, i.e. when the O_2 content rises the CO content decreases, and vice versa.

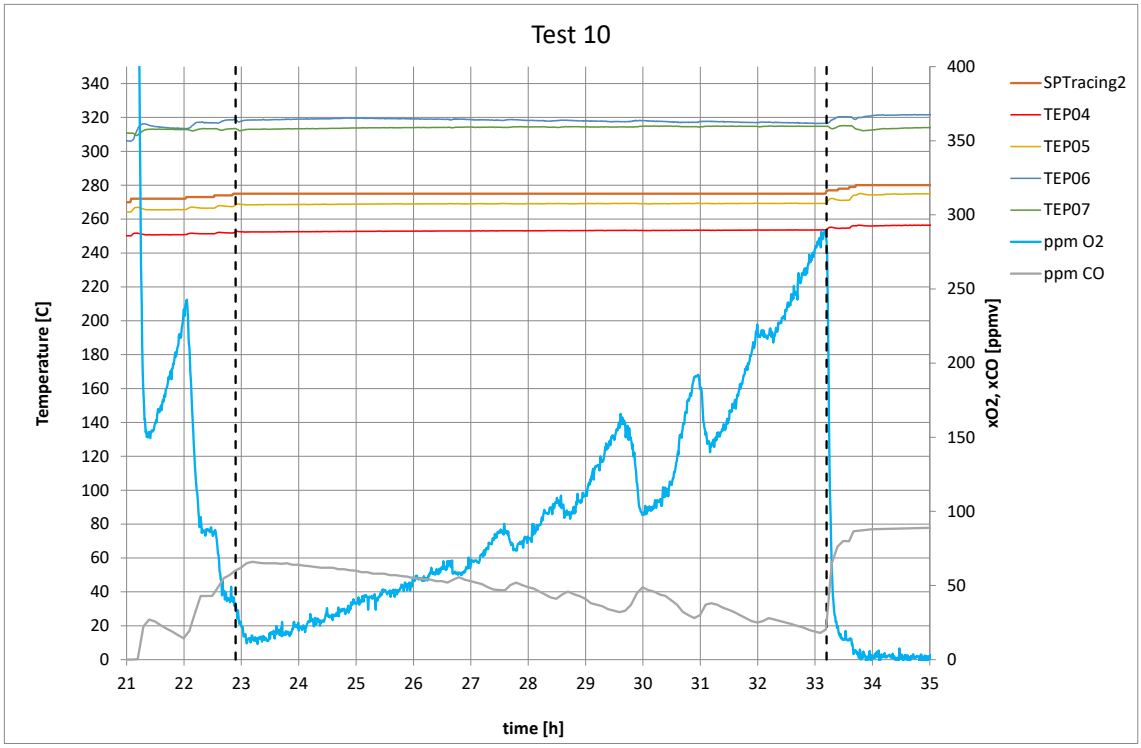


Figure 7: Temperature and concentration profile development during Test 10 (base case, Catalyst B).

It was remarkable anyway that catalyst B showed formation of CO even before the O_2 conversion was complete. CO-levels of 50-100 ppmv were observed before O_2 conversion was complete. Catalyst A also showed some CO formation, but only above the temperature at which the O_2 conversion was complete; moreover CO was produced at lower levels than with catalyst B. Formation of CO can be explained by the occurrence of the steam reforming reaction:

$CH_4 + H_2O = CO + 3H_2$

It might be expected that formed CO will quickly oxidize to CO_2 if O_2 is still present (same for H_2, but this was not measured). But apparently CO and O_2 can co-exist on catalyst B. When CO is being formed it was also noted that the CO_2 concentration increased somewhat (not shown), which may be explained by the water-gas shift reaction:

$CO + H_2O = CO_2 + H_2$

The measured CO_2 concentration was (slightly) higher than can be explained by methane oxidation only, which points to a reaction involving water.

3.3 Results for catalyst A

3.3.1 Variation of the flow rate

In Test 6 the flow was doubled compared to the base case (8 slm vs. 4 slm). This resulted in somewhat higher temperatures to reach full conversion, both when increasing the temperature (by 6ºC) and by decreasing the temperature (by 14ºC). This is more or less expected.

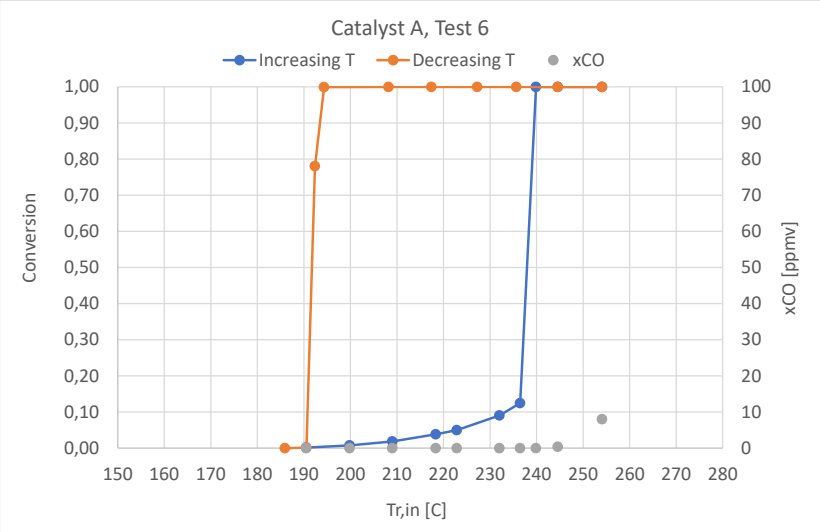


Figure 8: Results for Test 6 (higher flowrate than the base case)

In Test 7 the flow was also doubled compared to the base case (8 slm vs. 4 slm); on top of this the pressure was also increased to 1.8 bar(a), compared to 1.2 bar(a) for the base case. This resulted in somewhat higher temperatures to reach full conversion, both when increasing the temperature (by 6ºC) and by decreasing the temperature (by 14ºC). This effect was about the same as for Test 6, in which only the flow was doubled; hence the pressure increase added no measurable extra effect.

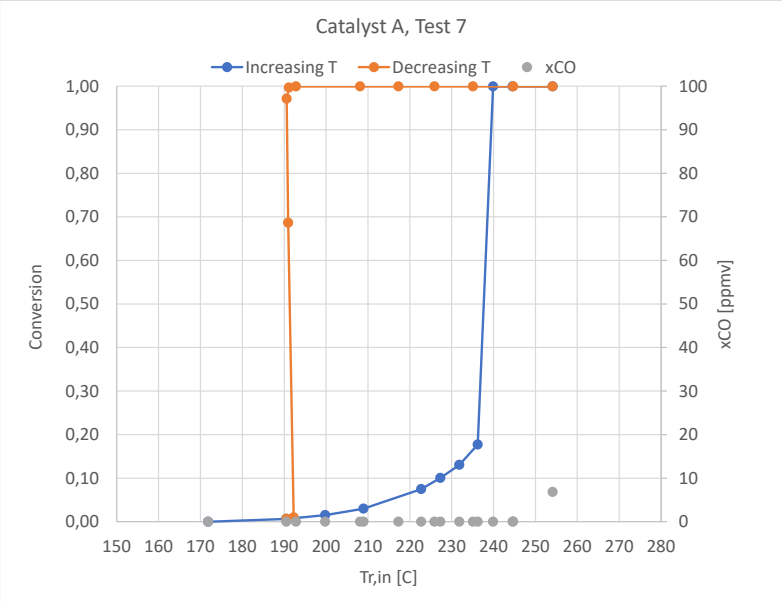


Figure 9: Results for Test 7 (higher flowrate and higher pressure than the base case)

3.3.2 *Variation of the oxygen content*

In Test 2 the O_2 concentration in the feed was halved compared to the base case (0.5% vs. 1.0% in the base case). During temperature increase full conversion was now reached at a lower temperature than in the base case (by 16ºC), however during temperature decrease the lowest temperature with full conversion was now higher (by 11ºC). The latter might be explained by the lower amount of heat produced by the reaction (with a consequently lower temperature increase across the reactor).

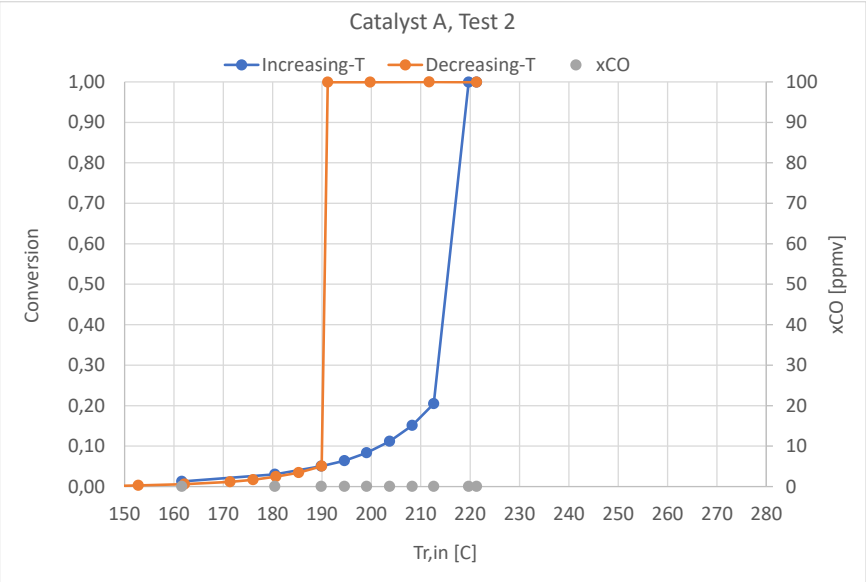


Figure 10: Results for Test 2 (50% lower oxygen content of the feed than the base case)

In test 1, again the oxygen level was halved compared to the base case test, and on top of this the water in the feed was set to 0 (compared to a feed water content 0.25% in the base case). At increasing temperature the temperature for full conversion was now 24ºC lower than for the base case, mainly due to the lower O_2 content of the feed. The lower water content causes a bigger temperature difference than test 2, which showed a 16ºC lower temperature. On the other hand, at decreasing temperature the lowest temperature for full conversion was now 11ºC higher than in the base, which was about the same effect as in test 2. Apparently the absence of water had no detectable effect during temperature decrease.

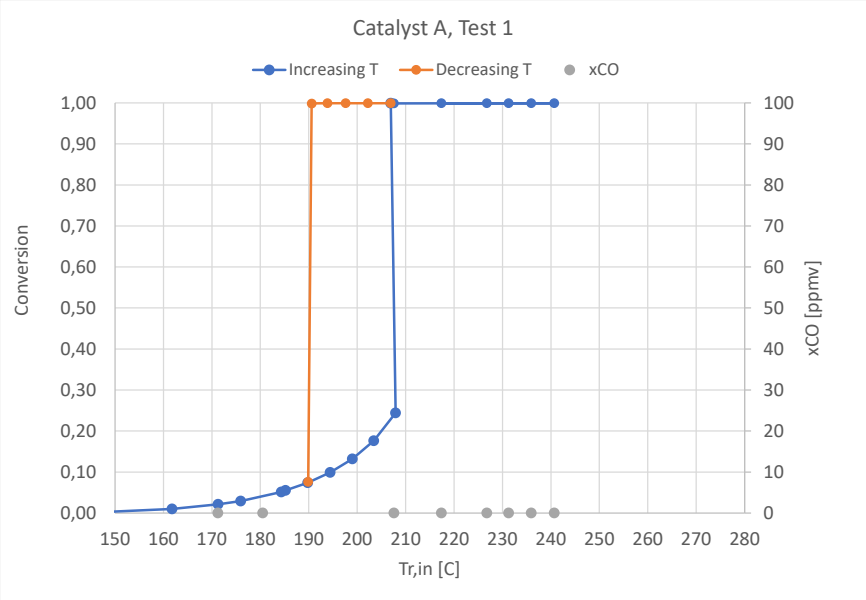


Figure 11: Results for Test 1 (50% lower oxygen content than the base case, no water in the feed)

3.3.3 *Influence of pressure*

In test 5 the pressure was increased compared to the base case (1.8 vs. 1.2 bara). There was hardly a measurable effect on the temperature at which full conversion was achieved. The pressure difference was small.

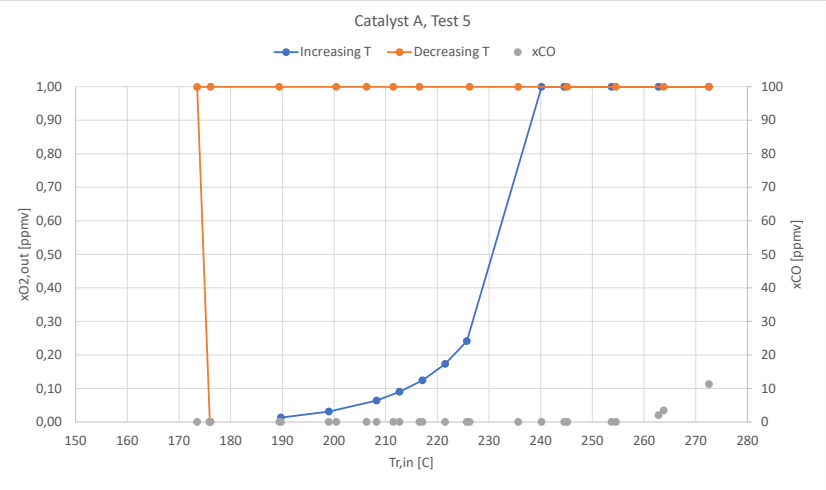


Figure 12: Results for Test 5 (pressure 1.8 bara compared to 1.2 bara in the base case)

3.3.4 *Variation of the water content*

In test 3 the H_2O content of the feed was set to 0, compared to 0.25% in the base case test. The test was interrupted by an alarm in the laboratory, and was restarted; hence more than one increasing and decreasing temperature curve were measured, which gives an (unplanned) insight of reproducibility of the tests. All in all the temperatures for reaching full conversion were just slightly lower than in the base case test (by about 2ºC for increasing and by about 4ºC for

decreasing temperature). This is probably explained by adsorption of H_2O on the catalyst surface, occupying active sites that are no longer available for the oxidation reaction. However the amount of water in the feed is quite small compared to the amount of water produced by the reaction; hence the effect of water in the feed is not very outspoken.

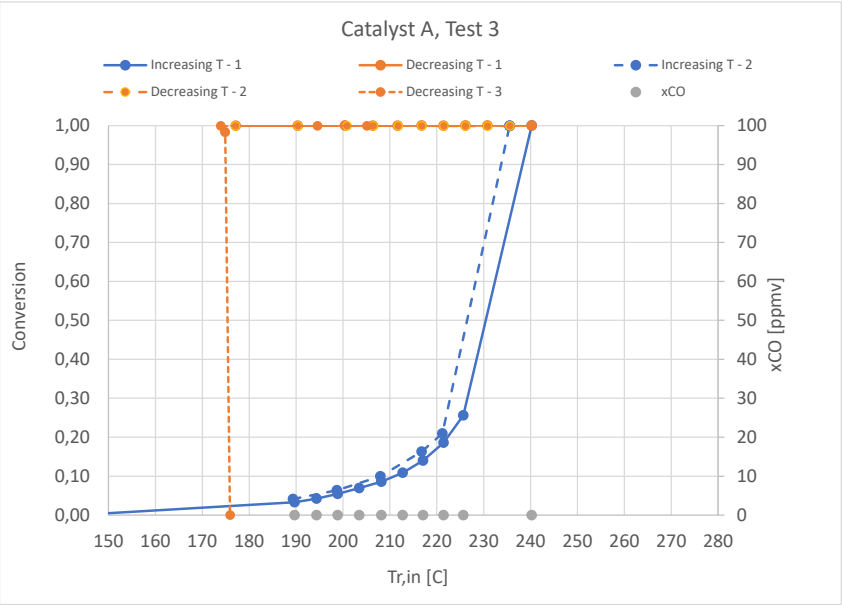


Figure 13: Results for Test 3 (no water in the feed, compared to 0.25% water in the base case)

3.3.5 Overview of test results for catalyst A

In table 3, the impact of the variation of test conditions on the minimum temperature for full conversion in the tests with catalyst A is summarized. The biggest impacts appear to be the oxygen content and the flow rate.

Table 3: Lowest temperature for full conversion for Catalyst A

Test	Parameters varied	Increasing temperature		Decreasing temperature		Lowest x_{O2} in product gas [ppmv]
		Temperature for 100% conversion	Difference with base case	Temperature for 100% conversion	Difference with base case	
1	$x_{O2}\downarrow$, $x_{H2O}\downarrow$	208ºC	-24 ºC	190 ºC	+11 ºC	3
2	$x_{O2}\downarrow$	216 ºC	-16 ºC	190 ºC	+11 ºC	3
3	$x_{H2O}\downarrow$	230 ºC	-2 ºC	175 ºC	-4 ºC	2
4	Base case	232 ºC	-	179 ºC		2
5	P↑	233 ºC	+1 ºC	175 ºC	-4 ºC	4
6	Flow↑	238 ºC	+6 ºC	193 ºC	+14 ºC	5
7	Flow↑, P↑	238 ºC	+6 ºC	193 ºC	+14 ºC	5

The final column of Table 3 shows the minimum O2 content measured at the top of the conversion curves, to give an impression of the achievable levels. They are all well below 10 ppmv.

3.4 Results for catalyst B

3.4.1 Variation of the flow rate

In Test 12 the flow was doubled compared to the base case (8 slm vs. 4 slm). This resulted in somewhat higher temperatures to reach full conversion, both when increasing the temperature (by 16ºC) and by decreasing the temperature (by 17ºC). This is more or less expected.

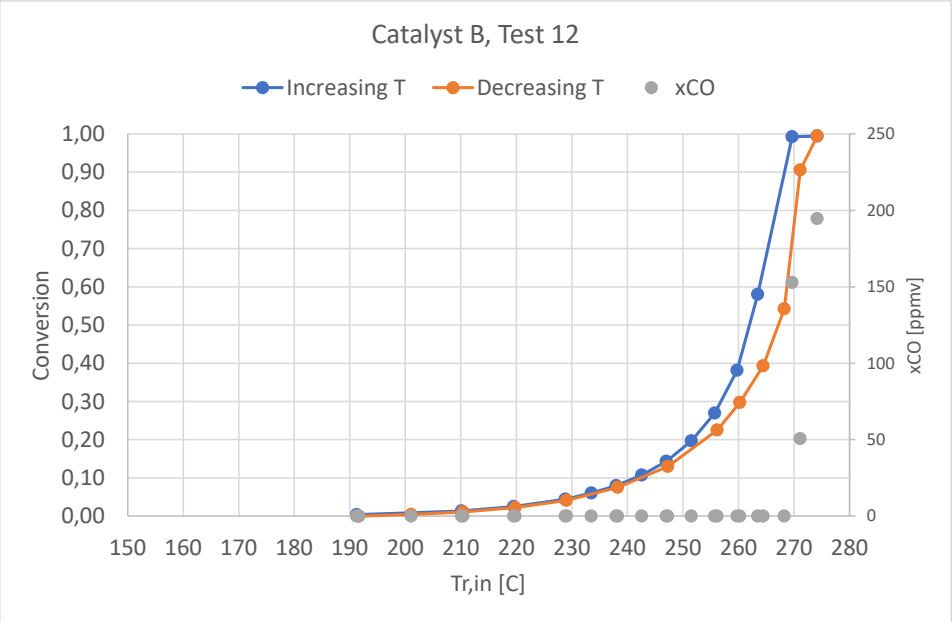


Figure 14: Results for Test 12 (higher flowrate than the base case)

It is noticed that the maximum CO concentration was quite high at about 200 ppm. This was most likely related to the high temperatures in this test.

In Test 13 the flow was doubled compared to the base case (8 slm vs. 4 slm), and in addition the pressure was increased as well to 1.8 bar(a) (compared to 1.2 bar(a) in the base case). This resulted in higher temperatures to reach full conversion, both when increasing the temperature and by decreasing the temperature (both by 17ºC). This was the same as for Test 612 in which only the flow was doubled; hence the pressure increase added no measurable extra effect.

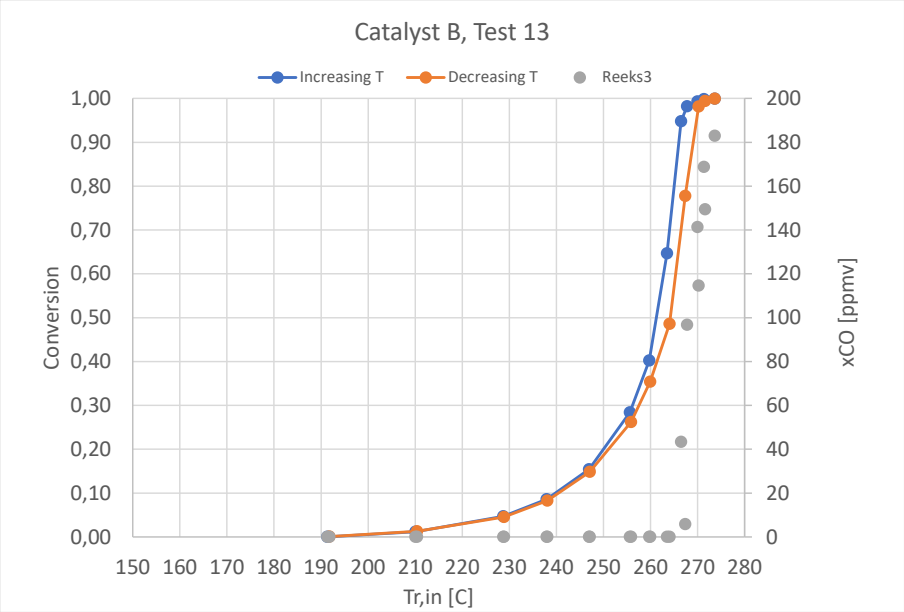


Figure 15: Results for Test 13 (higher flowrate and higher pressure than the base case)

3.4.2 Variation of the oxygen content

In Test 8 the O_2 concentration in the feed was halved compared to the base case (0.5% vs. 1.0% in the base case). Full conversion was now reached at a lower temperature than in the base case: by 6ºC and 5ºC at increasing and decreasing temperatures respectively.

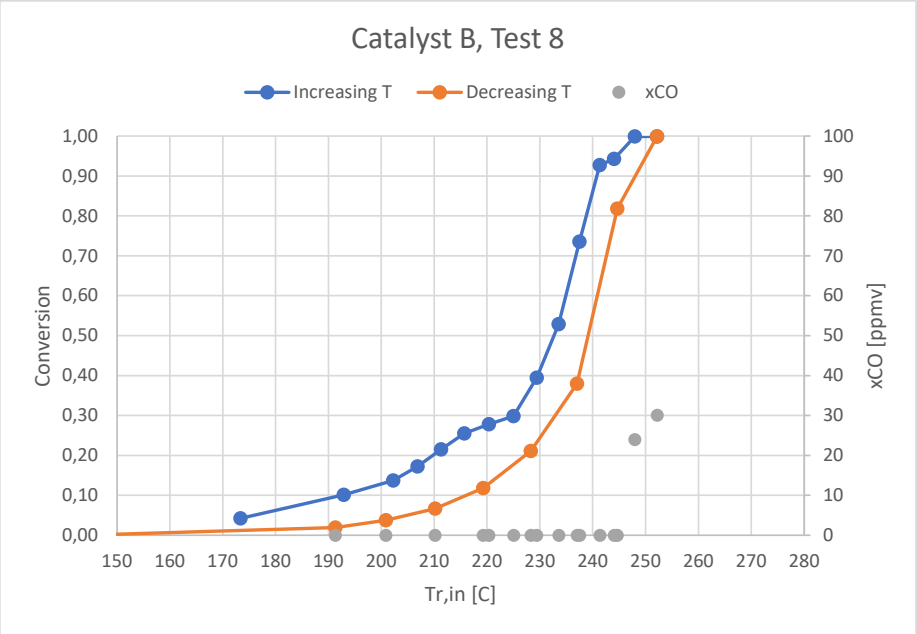


Figure 16: Results for Test 8 (lower feed oxygen content than the base case)

3.4.3 *Variation of the pressure*

In test 11 the pressure was increased compared to the base case (1.8 vs. 1.2 bara). There was hardly a measurable effect on the temperature at which full conversion was achieved. The pressure difference was small.

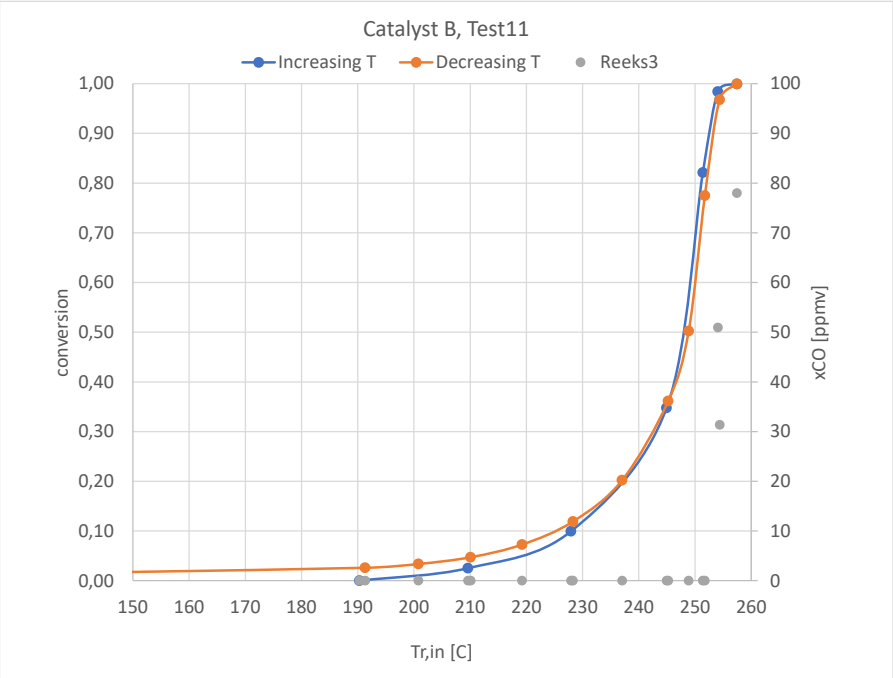


Figure 16: Results for Test 11 (pressure 1.8 bara compared to 1.2 bara in the base case)

3.4.4 *Variation of the water content*

In test 9 the H_2O content of the feed was set to 0, compared to 0.25% in the base case test. The temperatures for reaching full conversion were equal to the ones in the base case test.

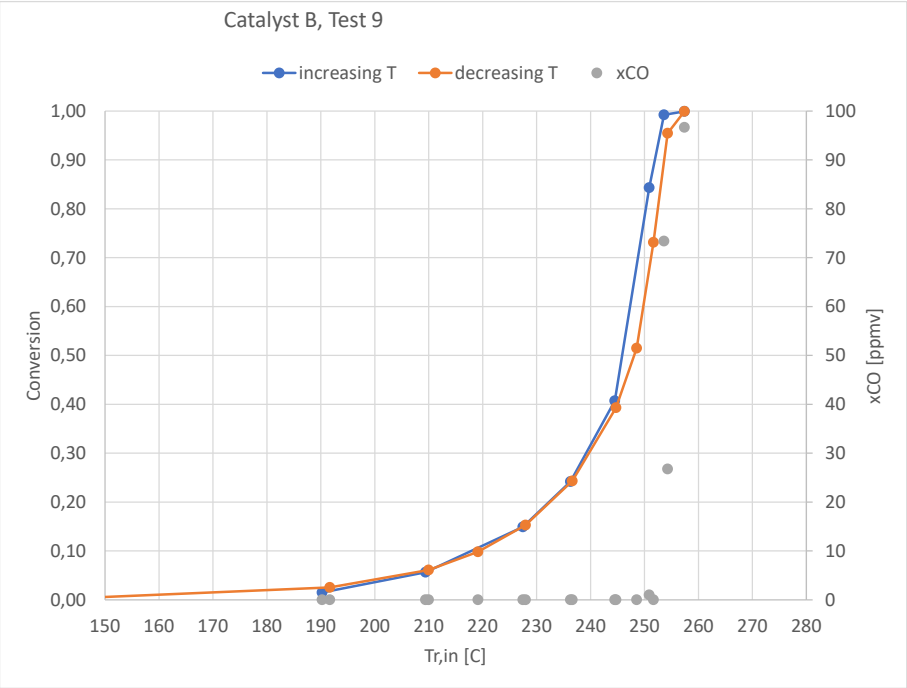


Figure 17: Results for Test 9 (no water in the feed, compared to 0.25% water in the base case)

3.4.5 Overview of test results for catalyst B

In table 4, the impact of the variation of test conditions on the minimum temperature for full conversion in the tests with catalyst B is summarized. The biggest impacts appear to be the oxygen content and the flow rate.

Table 4: Lowest temperature for full conversion for Catalyst B

Test	Parameters varied	Increasing temperature		Decreasing temperature		Lowest x_{O2} in product gas [ppmv]
		Temperature for 100% conversion	Difference with base case	Temperature for 100% conversion	Difference with base case	
8	$x_{O2}\downarrow$	248 °C	-6 °C	252 °C	-5 °C	4
9	$x_{H2O}\downarrow$	254 °C	0 °C	257 °C	0 °C	7
10	Base case	254 °C	-	257 °C	-	5
11	P↑	257 °C	+3 °C	257 °C	0 °C	7
12	Flow↑	270 °C	+16 °C	274 °C	+17 °C	52
13	Flow↑, P↑	271 °C	+17 °C	274 °C	+17 °C	5

The final column of Table 4 shows the minimum O_2 content measured at the top of the conversion curves, to give an impression of the achievable levels. In most tests the O_2 levels are below 10 ppmv, except for test 12 in which the conversion was actually not high enough (the temperature should have been increased slightly more). The measured level occurred after a testing period of 14 h; initially the O_2 level was below 10 ppmv but over time it had slowly increased (illustrating again the less stable behaviour of this catalyst). CO levels were high, i.e. over 200 ppm.

4 Conclusions and recommendations

In this work, two commercial catalysts have been tested for their oxygen removal capacity from biogas in a laboratory setup. Conversions were measured while the temperature was first increased in steps, and then decreased in steps. The minimum temperature at which full conversion is reached is a measure for the catalyst activity. The results showed that O_2 conversions over 99.9% can be reached by operating at a suitable temperature.

Catalyst A showed a wide hysteresis in the activity between heating up and cooling down, while catalyst B did not show such behaviour. Catalyst A turned out to be more active than catalyst B: on the whole, catalyst B requires 30-40ºC higher temperatures to achieve full conversion compared to catalyst A, based on the curve for increasing temperature, and requires even 60-80ºC higher temperatures based on the curve for decreasing temperature.

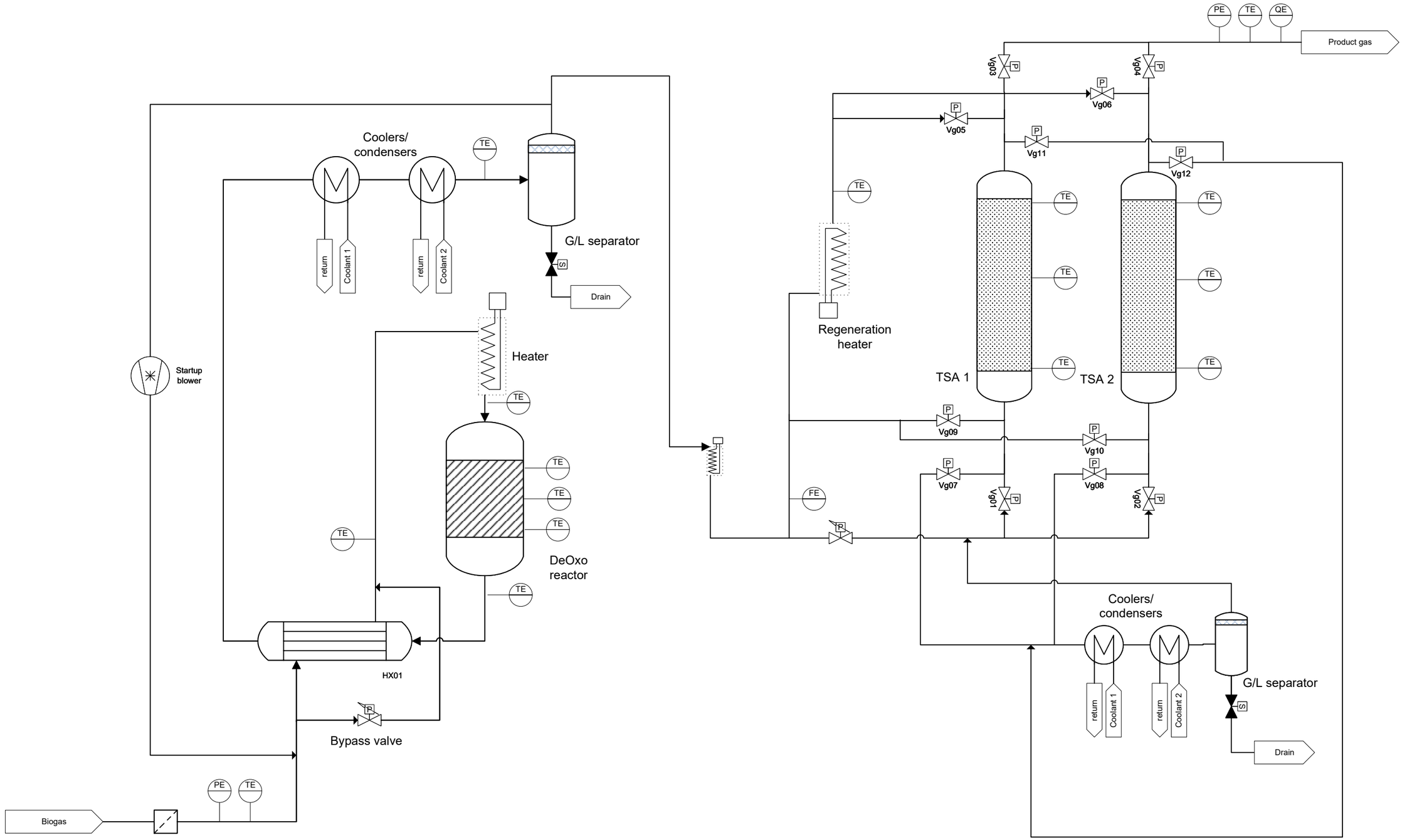
It was observed that catalyst B produces considerable amounts of CO, even when oxygen conversion is not complete. Catalyst A also produces some CO, but far less than catalyst B, and only when the oxygen conversion is complete. In addition, for catalyst B it was often observed that the oxygen conversion was not stable. The variation in the O_2 levels in the product gas seems to be related to simultaneous variations in CO production.

The impact of several process conditions was determined in a number of tests. As expected a higher flow rate requires a higher temperature to achieve full conversion, whereas a lower oxygen content of the feed allows a lower temperature to achieve full conversion. The impact of changing these factors is comparable for both catalysts, although possibly the effect of flow is somewhat more outspoken for catalyst B and the effect of the oxygen content is somewhat more outspoken for catalyst A. Variation of the water content of the feed did not show a significant effect; this can be understood from the fact that the amount of water added to the feed was not very high compared to the amount of water produced by the chemical reactions. The pressure was also varied, but the variation was only very small and did not result in a measurable effect.

It is recommended to pursue further development work based on the application of catalyst A. The test results indicate that a reaction design based on a GHSV in the range 5000-10000 h^{-1} should be feasible; this determines the required catalyst bed volume for a given feed flow rate.

The impact of pressure could not be well established in this work. However it may be expected that at high pressure the reaction rates will be higher, and it will be only easier to achieve high conversions. This is confirmed by communication with the manufacturer of the catalysts. In addition the manufacturer expressed the expectation that side reactions will be suppressed at higher pressures; but data are not available to the authors. It might be interesting to establish the effect of high pressure especially for catalyst B to see if the conversion would be more stable and CO production less.

Given the sometimes slow response of the conversion to changes in temperature a DeOxo-reactor requires careful control and monitoring of the product gas composition. Measurement of the development of the bed temperature profile may give a clue whether the reaction is slowing down or speeding up; but this should also be related to changes in the feed flow rate and composition.



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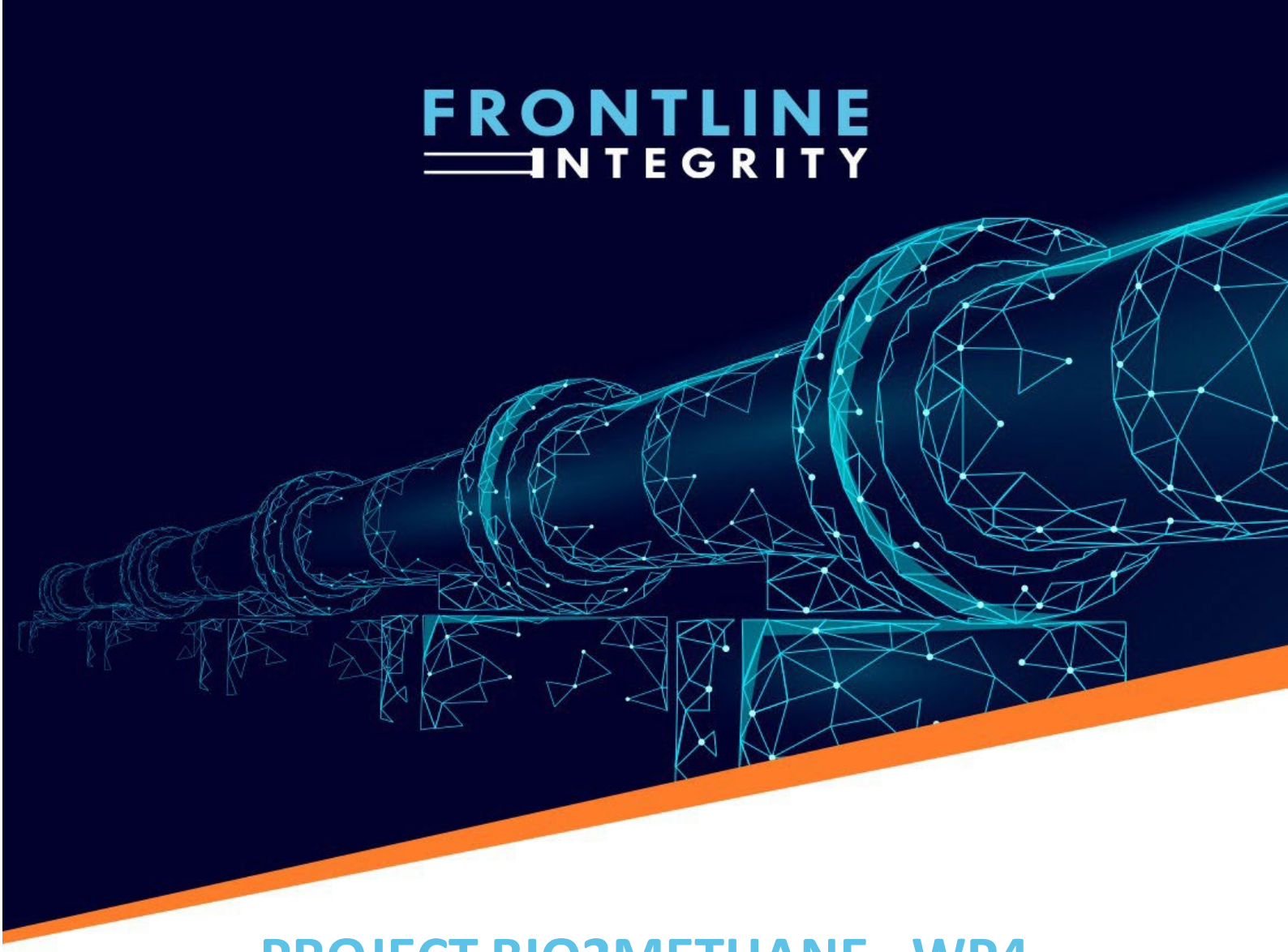

PROJECT BIO2METHANE - WP4 (CORROSION THREAT AND MANAGEMENT)

347-NGT-REP-001

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Disclaimer

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0	23/12/2026	Draft for Comment	SDR / RK	BK	SC

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Abbreviations

FRONTLINE INTEGRITY have used the following abbreviations in this report:

Term	Definition
BS	British Standard
CBA	Cost Benefit Analysis
CIPS	Close Interval Potential Survey
COMAH	Control of Major Accident Hazard
GSMR	Gas Safety Management Regulations
HSE	Health and Safety Executive
IRIS	Internal rotary inspection system
NEA	Network Entry Agreement
NGT	National Gas Transmission
NTS	National Transmission System
PRCI	Pipeline Research Council International
PSR	Pipeline Safety Regulations
PSSR	Pressure Systems Safety Regulations
RBI	Risk Based Inspection
UESO	Underground Energy Storage Operators (Limited)
US	United States
UTM	Ultrasonic Thickness Measurement
UK	United Kingdom
VCI	Volatile Corrosion Inhibitors
WP	Work Pack

Executive Summary

National Gas Transmission (NGT) is undertaking the BiO2Methane project (GTX/06888) which focusses on the removal of oxygen from biomethane so that it can be stored in underground storage sites without corrosion compromising the integrity of the storage and associated facilities. Frontline Integrity has been appointed to assess the corrosion threat and management.

This report is associated with work pack 4 which includes:

- The assessment of the corrosion threat considering increased oxygen levels
- Review of possible Corrosion Mitigation Options
- Future Risk Management Considerations

Conclusions

Corrosion Threat Review

An oxygen level in excess of 10 ppm is likely to elevate the threat of internal corrosion. The predicted corrosion threat increases as product temperature decreases. As an example, 0.2 mol % (2000 ppm) at 98 bar pressure is predicted to result in a corrosion rate of 0.3 mm per year at 20° C.

Limiting the oxygen content to 10 ppm is likely to limit the corrosion threat to levels below 0.05 mm per year. However, the presence of carbon dioxide may contribute to elevated corrosion pitting rates.

The most susceptible areas to internal corrosion have been assessed as:

1. Subsurface pipelines from wellheads into the storage cavern: Depending on operating conditions, condensation may occur. Wet gas conditions may occur in the form of a water stream along the walls of the pipe in annular mist flow.
2. Above ground pipework: Depending on operating conditions, there may be water drop out before dehydration which would result in an elevated corrosion threat at low points.

The corrosion assessment completed in this analysis and the conclusions above are based on several scenarios which were deemed representative of the data provided by all underground energy storage operators.

Internal Mitigation Review – Corrosion Inhibitor

This study has reviewed the corrosion inhibition options as possible mitigation. Our evaluation has concluded that options are limited and there is no ‘tried and tested’ solution in the market to reduce oxygen. Further work would be required to investigate and validate this as a suitable option.

Internal Corrion Mitigation Review – Internal Lining of Pipelines

This study has reviewed internal lining of in situ pipelines as possible mitigation option. Our evaluation has concluded that whilst options are available for short above ground sections, the in-situ lining of subsurface pipelines from the wellhead into the storage facilities was assessed as not currently feasible. Further work would be required to investigate and validate this as a suitable option.

Recommendations

Corrosion Threat and Risk Management

Considering the elevated threat of internal corrosion at increased oxygen levels, it is recommended that the following is considered per site:

- If not already in place, each of the underground energy storage operators should perform a site-specific corrosion assessment considering their specific pipeline design, historic and future operating envelope to assess the corrosion threat and confirm the most susceptible locations.
- Ensure suitable corrosion mitigation is installed to monitor or mitigate the threat ensuring that the oxygen content is below 10 ppm as a minimum (e.g. the installation of oxygen removal equipment).
- Perform updates to the Safety Case and Written Scheme of examination, driven by careful review and revision of the relevant risk assessments. This may require more detailed sectionalising than before considering the areas of increased internal corrosion susceptibility.
- If not already in place, develop a long-term internal corrosion management plan to monitor the threat.
- Ensure all associated procedures, competency or certification requirements are reviewed and updated if required.

Suggestions for further Study:

Suitability and Feasibility: If deemed appropriate, perform further evaluation of the mitigation reviewed as part of this study (e.g. corrosion inhibition). This study could review what testing is required to further develop or validate the option, how could these be installed and identification of any uncertainties or risks.

Contingency Planning (site specific)**:** Investigation of the likely corrosion threat during an operational outage where the selected mitigation has failed (e.g. Oxygen removal equipment or inhibitor injection failure). What are the threats, what levels of oxygen would be likely, where would corrosion occur and how long can the system be operational without mitigation before it becomes a significant issue.

Risk Management (Site Specific): Perform a corrosion screening assessment across all storage sites to identify the sites likely to have seen the most internal corrosion to date or sites likely to be exposed to elevated corrosion rates in the future considering operating conditions.

In order to baseline the condition, review the feasibility of inspecting the subsurface pipework of the highest risk site with tethered or robotic in-line inspection technology to assess the current condition, confirm how much of the corrosion allowance has been consumed to date and enable routine monitoring during future operation. This would provide reassurance on the other storage sites and help to demonstrate regulatory compliance.

1 Introduction

National Gas Transmission (NGT) is undertaking the BiO2Methane project (GTX/06888) which focusses on the removal of oxygen from biomethane so that it can be stored in underground storage sites without corrosion compromising the integrity of the storage and associated facilities. Frontline Integrity has been appointed to assess the corrosion threat and management.

Project activities have been split up into the following Work Packs.

- WP1 – Project Management and Business Case. This work pack was led by NGT with contribution from other partners and included project management activities, as well as a cost benefit analyses.
- WP2 – Oxygen Removal. Undertaken by Hygear B.V. (Hygear) with inputs from UESO involved determining appropriate technology for oxygen removal that is both technically feasible and economically viable.
- WP3 – Integration with Underground Storage. This involved the development of conceptual designs for the integration of oxygen removal technologies with existing underground storage sites. The work was undertaken by Premtech (Ltd) with inputs from Hygear and UESO.
- WP4 – Corrosion Prevention Measures involved assessing of corrosion mitigation as well as evaluating corrosion risk and management. This is led and undertaken by Frontline Integrity with inputs from UESO.

1.1 Scope of Work

In addition to assessing the corrosion risk associated with increased oxygen content, Frontline's scope of work on the project includes assessing possible mitigations and improvements to corrosion management practices. These activities are included in WP4 and are described as follows:

- Review the likelihood of corrosion and calculate credible corrosion rates based on the project operating parameters and fluid composition.
- Evaluate potential mitigation measures, e.g. Inhibitor injection or coating application and estimate associated costs.
- Identify high risk areas and suggest improvements to existing inspection and maintenance protocols to support future corrosion management

2 System Boundaries and Asset Portfolio

The battery limits of this study are shown in red in Figure 2-1, which shows a schematic of the biomethane supply, transportation and storage system.

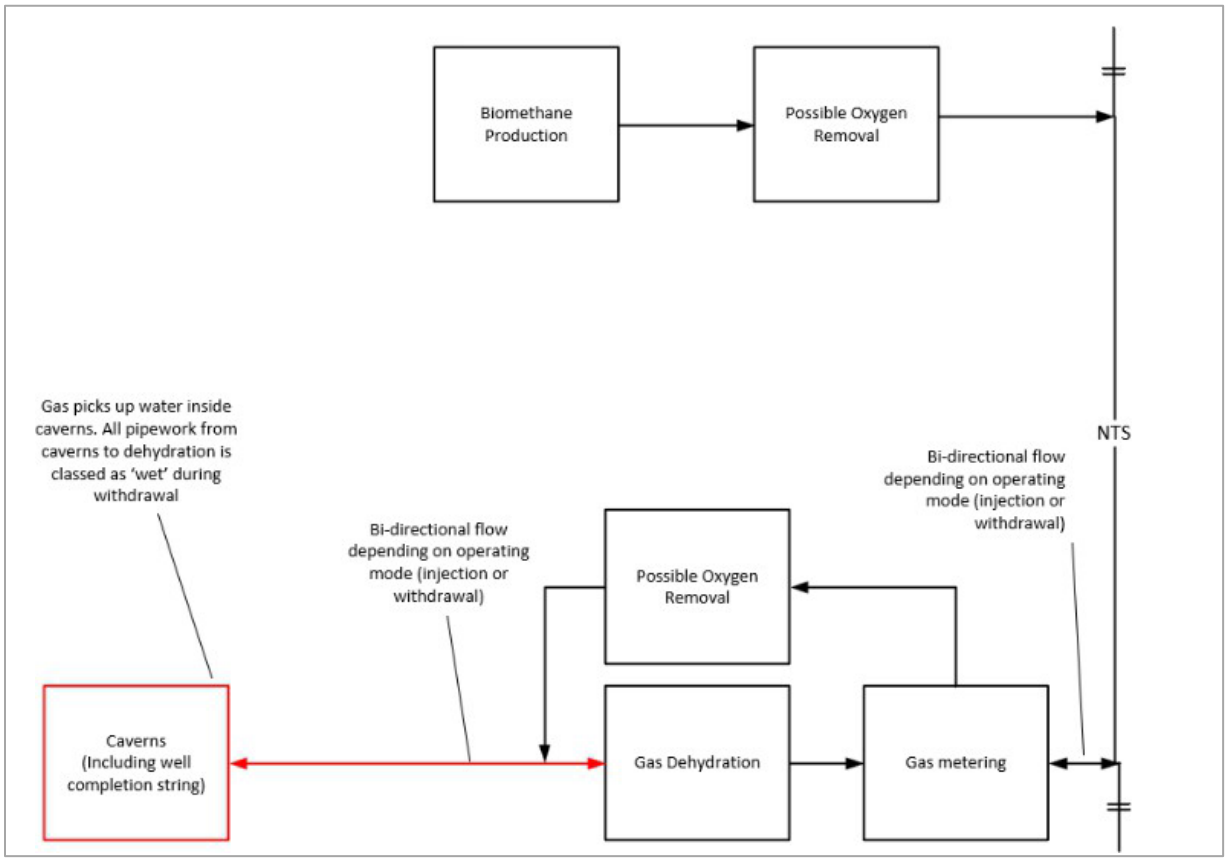

Figure 2-1 : Biomethane System Schematic

The study focusses on storage facilities in general rather than on a specific site.

Requests for information were issued to NGT, who subsequently distributed them to members of UESO. The objective was to establish the typical range of infrastructure involved and to understand current corrosion issues, along with mitigation and management approaches. In addition to storage data, details were also gathered on typical gas composition and the range of oxygen levels in the gas which was used for this assessment. These data for the basis of the findings presented in this document.

2.1 Asset Portfolio

2.1.1 Gas Storage

Underground gas storage facilities in the UK and Europe typically include man-made salt caverns, depleted oil and gas fields, and aquifers. Storage depths and capacities vary significantly; however, salt caverns are generally located at depths ranging from 500 m to 2000 m. These sites may be located either onshore or offshore. Due to the humid conditions within the facilities, gas stored in these environments typically becomes saturated.

2.1.2 Pipe Infrastructure

UESO provided responses from two storage operators. A summary of the information supplied by UESO, is summarised in Appendix A

The following types of pipework are typically found at storage sites:

- Bi-directional surface pipelines transporting gas between wellheads and processing facilities prior to connection with the National Transmission Pipeline System. The number, diameter and length of these pipelines vary depending on the location and number of caverns, as well as the design flow rate.
- Subsurface vertical tubing (completion strings) installed from the wellheads into the storage caverns, used for gas injection and withdrawal. The length of the pipelines depends on cavern depth, with typical reported lengths ranging from 500 m to 1800 m.

2.1.3 Asset Summary

The asset portfolio includes

- Bi-directional pipelines connecting wellheads and gas dehydration facilities at storage sites before connections to the National Transmission Pipeline System.
- Subsurface vertical pipelines from wellheads down into the storage caverns (completion strings).
- Associated fittings and components such as valves, pressure & temperature tapings, wellheads and impressed current cathodic protection systems.

2.1.4 Derivied Infrastructure

Based on the information provided, infrastructure scenarios were developed for analytical purposes to support this study. These scenarios do not replicate actual site configurations but are derived from those data provided. The scenarios were broadly categorised according to storage depth and pressure rating.

- Sites with approximately 500 m subsurface pipe depths and a pressure rating of around 100 barg.
- Sites with approximately 1700m subsurface pipe depths and a pressure rating of around 280 barg.

The typical pipeline scenarios that were derived are shown below in Table 2-1

Parameter	Case 1: Subsurface into Well	Case 2: Subsurface into Well	Case 3: Surface wellhead to plant	Case 4: Surface wellhead to plant	Case 5: Surface wellhead to plant
Length	[redacted]				
Diameter(s)					
Material types, grades & wall thickness					
Operating Pressure					
Design Pressure					
Operating Temperature					
Design Temperature (Min / Max)					
Volume flow rate					
Velocity					
Equipment Details					
Buried or above ground					

Table 2-1: Derived Infrastructure for Study

3 Corrosion Overview and Derived Gas Composition

3.1 Introduction

The transportation and blending of biomethane into gas transported by the NTS may result in an increase in the oxygen content. Biomethane producers face significant challenges in meeting the current low oxygen content specification arising from their production process. To address this, NGT is considering applying for an exemption from the HSE to deviate from the existing transmission network standards to allow up to 1 mol% oxygen to be transported in the pipeline network.

Since gas in the transmission system is typically dry, the increased oxygen levels would not, under normal circumstances, create a corrosive environment. However, specific risks arise in relation to underground storage facilities.

The following sections outline current transmission standards and their implications for biomethane blending.

3.2 Oxygen Content and Transmission Standards

Current NTS standards require very low oxygen levels, which are difficult for biomethane producers to achieve consistently. Granting the proposed exemption would enable biomethane to be blended into the NTS despite its higher oxygen content, without compromising compliance with other transmission specifications, since biomethane already meets all other required specifications.

While exemptions may resolve blending challenges, they introduce new considerations for underground storage, discussed next.

3.3 Underground Storage Impact

Gas from the NTS is stored in underground facilities that are inherently wet environments. Although extracted gas is dried before reintroduction to the NTS, the pipework infrastructure used to transport gas from storage facilities to the process and dehydration facilities, handles wet gas. The combination of increased oxygen content (due to biomethane blending) in combination with wet gas conditions immediately downstream of storage facilities introduces a corrosion risk that needs to be addressed.

To assess these risks, it is essential to review gas composition and oxygen limits as detailed in Section 3.4.

3.4 Gas Composition

3.4.1 Gas Composition Limits

For reference, gas composition limits as per GSMR requirements and a typical NGT Network Entry Agreement (NEA) are provided in Table 3-1.

Description	Content Limits	
	GSMR	NGT (Typical NEA Agreement (Note 1))
Hydrogen Sulphide	≤5 mg/m^3	≤5 mg/m^3
Total Sulphur	≤50 mg/m^3	≤50 mg/m^3
Hydrogen	≤0.1% (molar)	≤0.1% (molar)
Oxygen (Note 2)	≤0.2% (molar) ≤ 38 Barg	≤0.001% (molar)
	≤1.0% (molar) > 38 Barg	
CO_2	-	≤2.5% (molar)
Nitrogen	-	≤5.0% (molar)
Inerts (Note 3)	-	≤7.0%
Impurities / Contaminants	Note 4	Note 4
Hydrocarbon due point	Note 5	<-2°C @ 85 Bar
Water due point	Note 5	<-10°C @ 85 Bar
Relative density	≤0.700	-
Wobbe Index	≤51.41 MJ/m3	≤51.41 MJ/m3
	≥ 47.50 MJ/m3	≥ 47.20 MJ/m3

Table 3-1: GSMR & Example NGT NEA Gas Content Limitations

Notes:

1). Based on an example Network Entry Agreement (NEA) as included in DNV (2024) [1].

2). Oxygen content allowance was changed in 2025 to allow 1% in certain situations.

3). Inclusive of CO_2 and N

4). Shall not contain solid, liquid or gaseous material which may interfere with the integrity or operation of pipes or any gas appliance (with the meaning of regulation 2(1) of the 1998 regulations) which a consumer could reasonably be expected to operate.

5). Shall be at such levels that they do not interfere with the integrity or operation of pipes or any gas appliance (with the meaning of regulation 2(1) of the 1998 regulations) which a consumer could reasonably be expected to operate

3.4.2 Basis for Assessment

NGT requested that corrosion assessments use a gas composition representative of typical gas transported in the NTS, compliant with network specifications, but with varying oxygen content levels and assumed wet conditions.

In addition, typical biomethane compositions at production and actual gas sample measurements taken at the inlet of the Humbly Grove storage sites were provided. However, for the purposes of this study, the standard NTS specification gas is considered to be the most appropriate reference.

3.4.3 Current Oxygen Limits

The current NTS oxygen limit for transmission pipelines operating above 38 barg is 0.2% (2000 ppm), however in practice, actual oxygen levels in the network are reported to be significantly lower, typically in the range of 10 – 100 ppm [1].

3.4.4 Oxygen Scenarios Considered

For the purposes of this study, three oxygen levels have been considered:

- 0.001% (10 ppm) – Reflecting the oxygen removal performance achieved by solutions developed as part of WP2 and WP3 (by others), which will reduce oxygen levels to less than 10 ppm. This level also aligns with the limit adopted for gas supplied to sensitive users in Europe, such as underground storage, as reported by DNV [1]
- 0.2% (2000 ppm) – Based on the current GSMR & NTS oxygen limit for transmission pipelines operating above 38 barg.
- 1% (10 000 ppm) – Representing the revised maximum oxygen limit currently under consideration by NGT to accommodate introduction the introduction of biomethane into transmission pipelines operating above 38 barg.

3.4.5 Derived Gas Composition

The gas compositions derived from existing GSMR limits for use in this study are provided in Table 3-2. Note, the composition has not been adjusted for water content as this varies depending on conditions.

Component	Oxygen Concentration (mol %)		
	0.001	0.2	1
Hydrocarbons	97,074	96,881	96,105
Carbon dioxide	1.87	1.783	1.769
Nitrogen	1.132	1.13	1,12
Water	Refer to Table 3-3		
Oxygen	0.001	0.2	1
Hydrogen Sulphide	Assumed zero		
Total Sulphur	Assumed zero		
Hydrogen	Assumed zero		

Table 3-2: Derived Gas Composition

3.4.6 Water Content Assumptions

The water content of the gas depends on the pressure and temperature within the storage cavern. For illustrative purposes, typical water content values have been provided for caverns at depths of 500 m and 1700 m.

Gas becomes wet after entering a cavern. For this study a cavern temperature of 55°C has been assumed based on available information. Surface pipelines are assumed to have an average temperature of 20°C although some above ground sections may experience temperatures below 0°C during winter. The calculated envisaged water content after storage is presented in Table 3-3 [2]

Pressure (barg)	Dissolved Water (mg/m3)		Water Dropout Volume (mg/m3)
	@ 55°C	@ 20°C	
98	1750	325	1425
286	900	175	725

Table 3-3: Water Content of Methane After Storage [2]

4 Assessment of Corrosion Threat Considering Increased Oxygen Levels

4.1 Overview

The following overview explains the causes of corrosion, with particular emphasis on the impact of water and oxygen concentration. Water facilitates but does not cause corrosion. The primary causes of corrosion are the presence of acid gases such as carbon dioxide and hydrogen sulphide, oxygen, sulphur and organic acids.

Corrosion may be general, though not uniform, or it may take the form of localized pitting. The acid gases and oxygen are all associated with pitting corrosion. Pitting caused by oxygen and hydrogen sulphide is particularly severe due to the formation of electrochemically active corrosion product layers on the surface of the steel. These layers provide partial protection to the underlying steel but also accelerate corrosion reactions at areas where steel is exposed.

In general, corrosion by oxygen will predominate, as also reported by DNV [1]. The relative severity of oxygen corrosion compared to carbon dioxide and hydrogen sulphide can be observed by reviewing corrosion data for these three gases, refer to Figure 4-1 . Oxygen corrosion causes significant corrosion at much lower concentrations than either carbon dioxide or hydrogen sulphide. Therefore, for the purposes of this study, only oxygen corrosion will be considered.

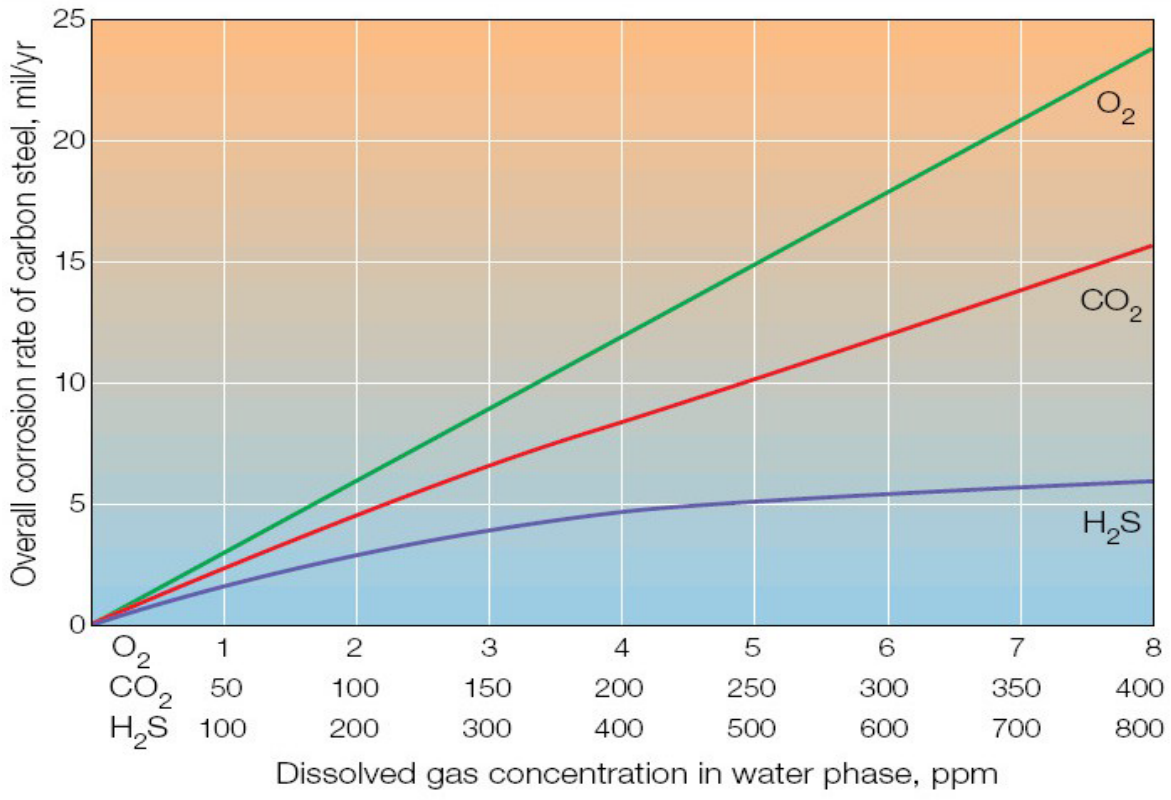

Figure 4-1: Corrosion Caused by Oxygen, Carbon Dioxide and Hydrogen Sulphide

The rate of corrosion is directly related to the concentrations of the oxygen (and other corrosive gases) dissolved in the water. At the gas storage facilities, this will be water that condenses when the warm gas from a storage cavern enters the cooler surface pipelines. Corrosion begins when the relative

humidity of the gas reaches approximately 50% and the rate increases with the amount of condensed water [3]. Higher corrosion rates will occur if sufficient water condenses to form a separate water phase within the pipeline.

Oxygen solubility in water is limited and depends on the concentration of oxygen in the gas phase, in addition to the pressure and temperature. The following sections describe how velocity, temperature and condensation rate influence the estimation of corrosion rates.

4.2 Effects of Velocity

There is limited literature available on the effect of gas velocity on condensing water films. However, flow morphology is a critical factor. Insights can be drawn from the many studies on water filled pipes and annular gas flow.

At low flow rates, the condensing water film is expected to drain slowly to the bottom of the pipeline whilst, at high flow rates, the morphology shifts to annular mist flow, where the water film streams along the pipeline surface driven by the gas flow. The impact of flow on corrosion differs between these two morphologies.

Available data indicates that, at low gas velocities, some form of stratified flow will dominate. This would be expected for pipelines with flow velocities of around 8 m/s (refer to typical case 3 of Table 2-1). As velocity increases, the water film on the upper surfaces of the pipeline becomes thinner, reducing the diffusion path and thereby increasing the corrosion rate. The water draining to the bottom of the pipeline will form a separate phase.

At a critical gas velocity, the flow morphology shifts to annular mist flow and the boundary layer thickness at the steel surface approaches an asymptote. This flow regime is most likely to occur in pipelines with velocities exceeding 12 m/s and up to 30m/s (refer to typical cases 4 and 5, Table 2-1). In this regime, water layer thickness stabilises resulting in a corrosion rate largely independent of gas velocity. Velocity may continue to have an effect by removing corrosion products from the surface, but the underlying corrosion rate would remain unaffected.

As pits develop, condensed water within the pits is largely shielded from the effects of gas velocity, reducing its influence on localised corrosion rates.

4.3 Effects of Temperature

Temperature has a significant influence on corrosion caused by oxygen. In an open system, the dissolved oxygen escapes from water as the temperature rises, reducing the rate of corrosion. In a closed system, such as a pressurized cavern or pipeline, oxygen cannot fully escape from the solution, and the corrosion rate continues to increase with rising temperature.

There is limited literature on how corrosion rates vary with temperature and almost none related to condensed water films. However, some indication can be drawn from a review of corrosion in water filled pipelines. Figure 4-2 shows that, for a given oxygen concentration, the corrosion rate approximately doubles or trebles when temperature is doubled.

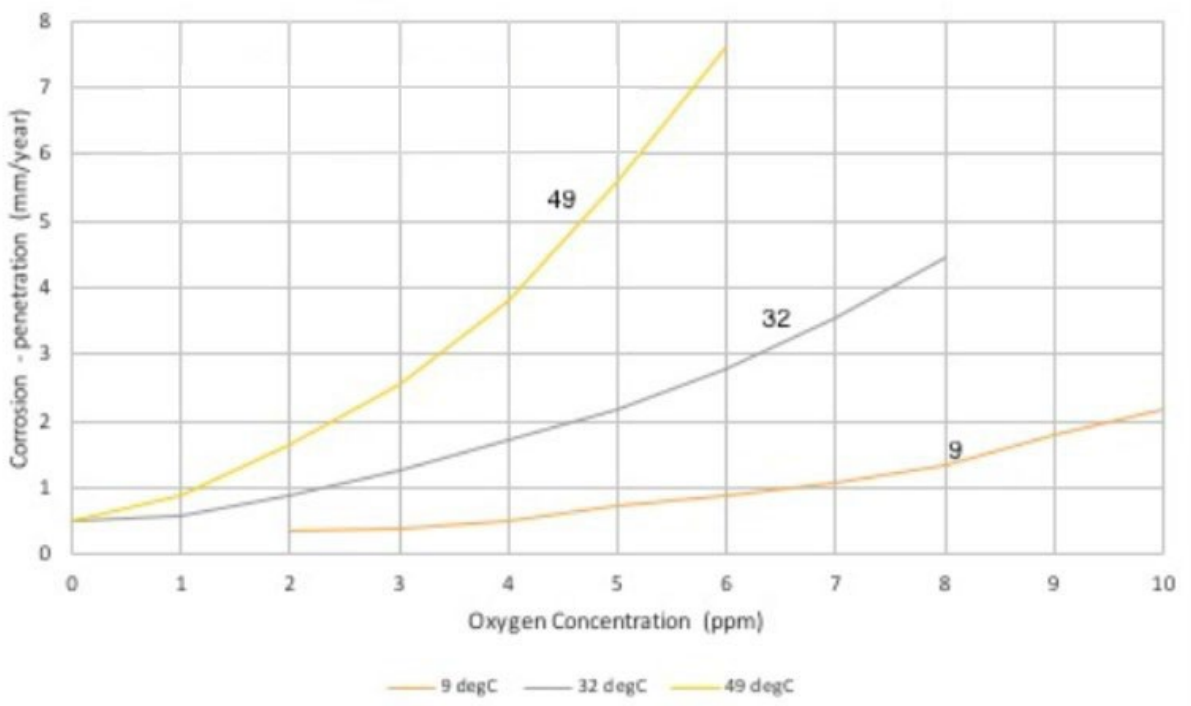

Figure 4-2: Impact of Temperature on Corrosion Rate

For the pipelines considered in this study, temperature effects are expected to be limited. Warm gas from the caverns will enter the surface pipeline and cool, with an estimated temperature difference of 40-50°C. The following section describes how temperature influences the solubility of oxygen in water.

4.4 Impact of Condensation Rates

highest closer to the source of the gas (wellhead) where gas temperatures and condensation rates are greatest.

Condensation rate depends on the temperature difference between the pipeline surface and the gas. Pipeline temperature is influenced by the rate of heat transfer between pipe and surrounding soil. Over distance, both gas temperatures and water condensation rate will gradually decline.

At low to modest gas flow velocities (below 12 m/s) and low water condensation rates, water tends to remain adherent to the pipeline surface, allowing corrosion to produce a build-up of ferrous ions in the water film. These ions react to form iron hydroxides and oxides that will remain on the pipe surface and create a diffusion barrier to oxygen, thereby reducing the corrosion rate.

At a high condensation rate the water film becomes unstable and flows down the pipe wall forming a discrete water layer at the bottom. In this case, the condensed water film on the pipe surface is replenished too quickly for ferrous ion concentrations to increase sufficiently for hydroxides and oxides to form on the pipe surface. As a result the corrosion rate remains high. There will be a difference in corrosion rate between the top and sides of the pipeline and the bottom of the pipeline. The critical condensation rate, based on the effect of condensation rate during carbonic acid corrosion, is 0.25 $gm/m^2/s$.[1]

1 Multifaceted Approaches for Controlling Top-of-the-Line Corrosion in Pipelines, W.W. Frenier & D, Wint, Oil and Gas Facilities, 67 – 80, June 2014

At high gas flow velocities (above 12 m/s), flow morphology may alter to annular mist flow. Water films condensing on the pipe surface will be driven axially along the pipe by the gas flow and some water will be dispersed into the gas flow as a mist. This occurs regardless of condensation rate. In this regime, corrosion rates are more uniform around the pipeline diameter due to the absence of a separated water phase.

The temperature decline profile is not known but to estimate the length of pipeline at risk of a higher corrosion rates downstream of the gas entry, a linear temperature decline to a soil temperature of 15°C was assumed. Temperature profiles were calculated for a decline to 15°C over a distance of 1000 m., although actual cooling distances may be longer.

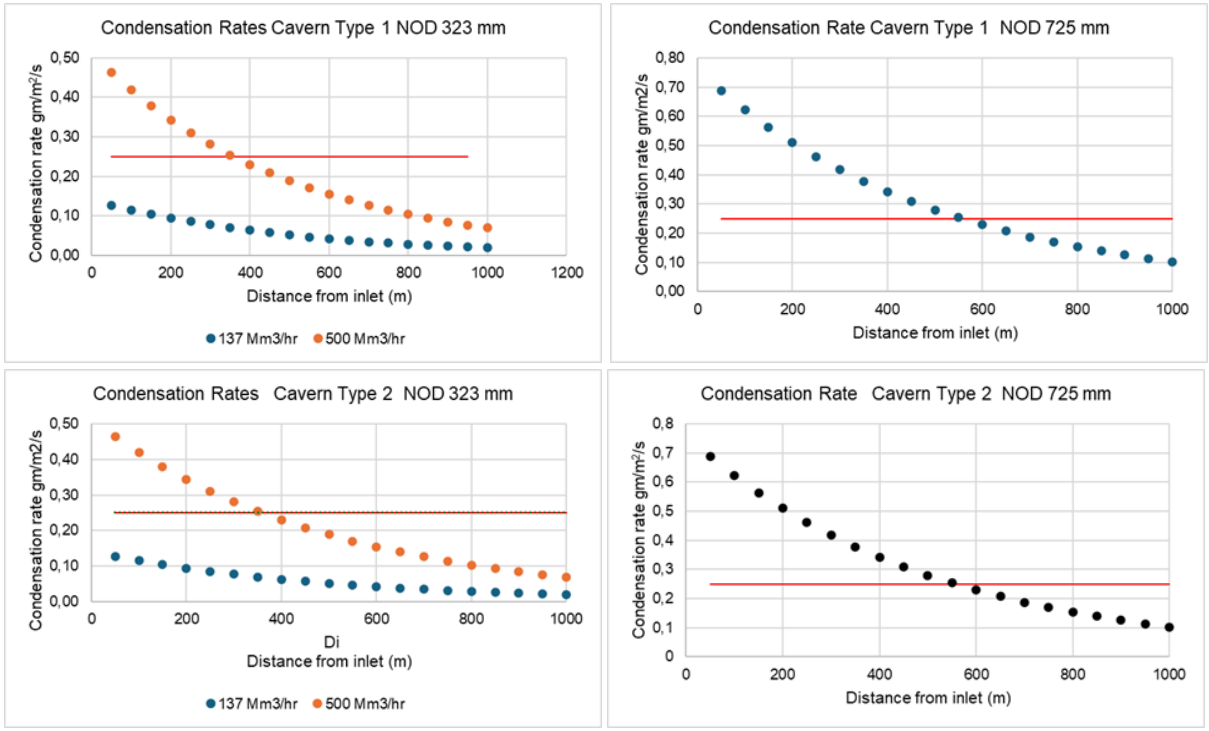

Figure 4-3: Decline of Condensation Rates with Distance for Various Cavern Scenarios

Sample condensation rate calculations indicate that water condensation is initially high but decrease as the pipe wall warms. Condensed water will either drain to the bottom of the pipeline, forming a separate phase, or be driven along the pipeline by the gas flow. Gas flow velocity determines which flow morphology will occurs.

If a thin stream of water collects at the bottom of the pipeline, there is a risk of localised corrosion at the 6 o'clock position, commonly referred to as 'Mesa corrosion'. This form of corrosion can lead to pipeline rupture if the affected area exceeds tolerable limits for internal pressure.

If the water flows axially along the pipeline, corrosion is likely to occur more uniformly across the entire inner surface of the pipeline.

4.5 Estimated Corrosion Rates

The rate of corrosion is directly related to the concentrations of corrosive gases dissolved in the water. In order to estimate corrosion rates for the project, the Tromans methodology was applied [4]. This methodology applies to pressures from 1 to 60 atmospheres and temperatures from 0 to 300°C.

Considering that gas will be stored and transported at pressures above 60 barg, oxygen values at higher pressures have been extrapolated from the 60 barg limit. These extrapolated values are unverified and should be considered as indicative only.

Dissolved oxygen levels for gas concentrations of 1 mol%, 0.2 mol% and 0.001 mol% have been calculated and are provided in Annexure B. The corresponding corrosion rates, which are directly related to dissolved oxygen concentration, are provided in Annexure C.

These tables were used to calculate the project corrosion rates as follows for pressure ratings provided for storage facilities of 98 barg and 300 barg, under temperature scenarios of 55˚C and 20˚C, as shown in Table 4-1 and Table 4-2 below.

	Corrosion Rate (mm/yr)	
	98 barg	300 barg
0.001	<0.05	<0.05
0.2	0.2	0.61
1	1	>1.5

Table 4-1: Predicted Corrosion Rates at 55°C

Oxygen Content (mol %)	Corrosion Rate (mm/yr)	
	98 barg	300 barg
0.001	<0.05	<0.05
0.2	0.3	0.95
1	>1.5	>1.5

Table 4-2: Predicted corrosion rates at 20˚C

When oxygen concentration is 1 mol%, dissolved oxygen range is 30-50 ppm and corrosion rates exceed 1.5 mm/year. At 0.2 mol%, the dissolved oxygen is between 2 and 8 ppm, with corrosion rates is in the range of 0.25-0.61 mm/year. At 0.001 mol%, the dissolved oxygen falls between 0.01 and 0.04 ppm and corrosion rates remain below 0.05 mm/year, although carbon dioxide will also contribute to corrosion.

It is not possible to predict corrosion rates with complete accuracy. However, it is evident that oxygen at concentrations above 10 ppmv have a major impact on corrosion rates and that severe pitting attack is likely at higher concentrations of oxygen.

In most pipeline sections, water condensation from the warm cavern gas will occur near the pipeline inlet. Depending on actual gas flowrate, water may form a separate water phase whilst or at high flow rates, stream along the pipe walls in annular mist flow.

4.6 Corrosion Assessment by DNV

DNV undertook a study on behalf of NGT to investigate the impact of increasing oxygen concentration limits to 1 mol % [1]. The study identified underground storage facilities as a sensitive user where higher oxygen content could potentially affect operations and the ability to deliver stored gas within the entry specifications.

One recommendation was that network flow studies should be undertaken for each biomethane injection application to determine whether the flow conditions would dilute biomethane oxygen content sufficiently before reaching any sensitive user.

The study focused on transmission within the NTS, where gas is dehydrated and concluded that corrosive conditions are unlikely to occur under normal circumstances. However, it highlighted that risks could arise if dehydration units become unavailable, increasing the likelihood of corrosion occurring needs to be addressed (such as unavailability of dehydration units).

Corrosion rates were calculated using formulas developed by the Pipeline Research Council International (PRCI), based on the following equation:

$$CR_{(mpy)} = 8.6988 + 9.856 \text{ x } 10^{-3}[O_2]^2 - 1.30865_{(pH)} + 4.934 \text{ x } 10^{-2}[CO_2][H_2S] - 4.8231 \text{ x } 10^{-5}[CO_2][O_2] - 2.372 \text{ x } 10^{-3}[H_2S][O_2] - 1.113 \text{ x } 10^{-3}[O_2]_{(pH)}$$

Where:

CR = Corrosion rate in millimetres per year (mpy)

O_2 = Oxygen concentration in Parts Per Million by Volume (ppmv)

CO_2 = Carbon dioxide partial pressure in Pound per Square Inch (psi)

H_2S = Haydrogen sulphide partial pressure in Pound per Square Inch (psi)

pH = Initial potential of hydrogen (pH) of the Solution

Calculations were performed undertaken for wet biomethane gas for varying carbon dioxide and oxygen levels assuming:

- Injection pressure of 94 barg.
- Presence of water and hydrogen sulphide (3.3 ppmv).
- Temperature of 16°C.

The study found that varying carbon dioxide concentrations (2.5%, 3.5% and 5%) had only a minor effect on corrosion rate, whereas rates were found to be more sensitive to changes in oxygen content. Figure 4-4 illustrates the effect of oxygen concentration on corrosion rate.

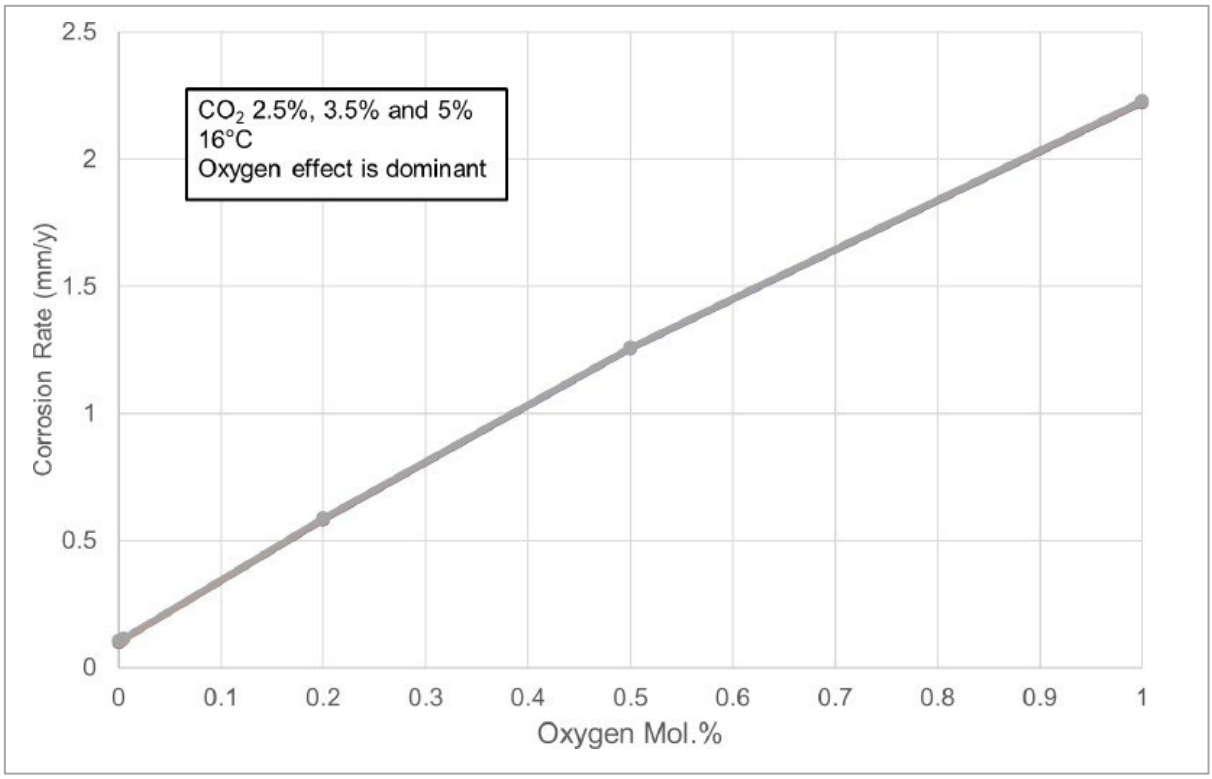

Figure 4-4: Effect of Oxygen Concentration on Corrosion Rate at 94 barg [1]

Table 4-3 below provides a comparison between the corrosion rates calculated as part of this project with those from the DNV study [1]. Note: the calculations were performed considering slightly dissimilar inputs. While the results are not the same, they correlate in order of magnitude, considering typical uncertainties associated with corrosion rate assessments.

Oxygen Content (mol %)	Corrosion Rate (mm/yr)	
	Frontline Integrity (98 barg / 20 °C)	DNV (94 barg / 16 °C)
0.001	<0.05	~0.1
0.2	0.3	0.6
1	>1.5	>2

Table 4-3: Comparison of Corrosion Rate Calculations

5 Corrosion Management Evaluation

Given the increased oxygen levels explored in this project, the corrosion threat is assessed as significant, as described above. This section explores options to manage the corrosion threat. Specifically, it evaluates the suitability of the following mitigation measures:

- Use of corrosion inhibitors
- Internal lining of pipelines

Descriptions of current corrosion management practices at storage sites and possible improvements are provided in the following sections of this document.

5.1 Corrosion inhibitors

5.1.1 Overview

Corrosion inhibitors are chemical additives used in the pipeline industry to prevent or reduce internal corrosion. They are typically applied in batches using pigs, or through continuous dosing systems, forming a protective barrier between the corrosive agents and the steel surface.

Inhibitors are used in gas and liquid pipelines but are generally more effective in liquid systems as the protective film is easier to maintain. In dry gas service, inhibitors face a number of challenges, such as:

- Uneven distribution in complex flow regimes,
- Inconsistent film formation
- Product incompatibility

The use of inhibitors adds to operational costs and requires regular monitoring and optimisation of dosing rates.

The following sections describe typical inhibitors used in the industry and their limitations in gas pipeline applications, followed by a feasibility assessment based on supplier engagement.

5.1.2 Typical Inhibitors

Common inhibitor types include:

- Film-forming amines / imidazolines / long-chain amines — These products adsorb at the metal/water interface and create a barrier film. Widely used in oil and gas production and sometimes applied for “top-of-line” corrosion in wet gas pipelines where a liquid film condenses on the upper pipe wall.
- Volatile Corrosion Inhibitors (VCIs) – Vapour-phase amine-type compounds that protect metal surfaces exposed to humid gas/vapour. VCIs have been trialled for top-of-line corrosion and equipment protection where liquid injection is difficult.
- Oxygen scavengers – Chemicals such as sulphite, hydrazine, organic hydroxylamine’s and, some patented volatile scavengers, remove dissolved oxygen from liquid phase. Common in closed water/boiler systems and can be used in wet pockets but rarely injected into gas pipelines.

5.1.3 Limitations associated with using Inhibitors

Several challenges limit the effectiveness of inhibitors in high oxygen gas environments:

- Continuous oxygen supply – When a continuous supply of oxygen is present, as is in conditions with high oxygen partial pressure for this project, any water film that forms will be highly aerated. Under these circumstances, oxygen reduction occurs rapidly, leading to elevated corrosion rates. Protective chemical films at the metal/water interface are compromised because oxygen is continuously replenished and flowing, turbulent flow can shear these films away.
- Film stability – Film forming inhibitors require a stable liquid film for persistent protection. In high-oxygen wet gas with thin, moving water films, protective films can be removed, diluted, or displaced. VCIs offer limited reach in a high velocity pipelines and have uncertain longevity.
- Chemical compatibility and downstream effects – Injecting chemicals such as amines, imidazolines or scavengers into a gas stream can cause downstream operational issues, including fouling, downstream contamination, catalyst poisoning, and gas quality violations. These risks create significant commercial and regulatory barriers, making continuous chemical injection into sales gas uncommon unless proven safe and contractually acceptable.
- Delivery challenges – Effective delivery and dosing requires reaching areas where water accumulates (low points, topside condensation zones). Achieving the correct local concentration in long pipelines is operationally challenging and batch treatments are generally impractical for long dry gas trunklines.
- Toxicity and regulation – Certain effective oxygen scavengers (e.g., hydrazine) are toxic and heavily regulated; others form by-products. This limits practical adoption for continuous injection into gas streams.

5.1.4 Feasibility Assessment and Supplier Feedback

The study examined the feasibility pf using corrosion inhibitors to mitigate corrosion caused by increased oxygen content in wet gas pipelines. Findings indicate that inhibitors are not widely used for this application due to several limitations discussed below. Instead, operators typically adopt oxygen removal and dehydration as primary corrosion control measures.

Specialist suppliers of corrosion inhibitor systems were consulted, but they were unable to offer a suitable solution for the specific application. Suppliers recommended the use of alternative corrosion management processes, such as removal of oxygen or dehydration. Consequently, no cost estimates are available for inhibitor-based solutions in this context.

Although inhibitor solutions for oxygen related corrosion exist specific localised problems, e.g. pigging bypasses, isolated wet pockets, equipment, no suitable solution was found as primary protection for long distance pipelines in the UK.

One product, CORR12949A (ChampionX), is currently used in the US to reduce corrosion rates on gas gathering lines with high oxygen content, where the inhibitor is applied in batches or through dosing systems. However, two of the raw materials used in the inhibitor are currently not available in the UK market and new product development will be required if this was to be deployed. Should inhibitor use still be considered as an alternative, further detailed studies will be required on the process (including further research and development) to allow use for the current application.

5.2 Internal Lining

Internal lining is an option to protect pipelines against internal corrosion. While internal linings are used on the NGT gas transmission network, they are typically applied under non-corrosive conditions to reduce internal friction and improve flow capacity.

This section evaluates options for internally lining existing pipelines at storage facilities, based on input from corrosion specialists and specialist contractors offering in-situ internal lining services. It is noted that the in-situ lining process is typically a complex process requiring detailed feasibility studies on a case-by-case basis.

5.2.1 Factory Applied Internal Lining (Pre-Installation)

Internal linings can be applied during the pipeline fabrication under controlled factory conditions, ensuring better quality control conditions. However, construction weld joints present challenges because the welding process damages internal coatings. Although techniques do exist to recoat joints, the cleaning and preparation of the subsurface prior to coating, as well as associated quality control, remain problematic.

Since this project focusses on existing storage facilities, factory applied internal linings would require pipeline replacement or re-installation, which is not considered practical and is therefore excluded from further consideration.

5.2.2 In-Situ Lining of Subsurface Pipelines (Completion Strings)

The feasibility of in-situ lining of subsurface pipelines from the wellhead into the storage facilities was assessed and found to be unfeasible.

These pipeline range from 500 m to 1700 m in length, which does not allow for the launching and receiving of batching pigs required for internal lining.

The alternative spray-applied lining was considered as an alternative. While some offer vertical spray solutions, the maximum reported application length is approximately 100 m. Additional limitations include:

- Cleaning requirements: Chemical cleaning and passivation are necessary before coating.
- Contamination risk: Cleaning chemicals and debris would enter the storage cavern, as there is no practical way to collect and remove them.

No successful case of in-situ lining of vertical completions strings was identified.

5.2.3 In-Situ Lining of Pipelines Between Wellheads and Processing Facilities

Options exist for in-situ lining of pipelines between wellheads and processing facilities, typically with diameters of DN 300, 500 and up to 750 and lengths ranging from a few hundred metres to 2-3 km.

Piggable pipelines can be lined using pig-applied epoxy / flow coat systems. Pigs are first used for cleaning the lines, typically foam, brush or scraper pigs, along with chemical cleaning and then coatings are applied either in batches between two pigs or spray applied. Pipelines are typically dried to the required dew point prior to the application of the lining.

Although feasible, practical considerations and risks that need to be considered, include:

- Achieving required level of cleanliness to ensure a durable coating application.

- Unknown buried pipeline conditions and risks associated with stuck equipment or sections of failed lining.
- Limited quality control and inability to perform acceptance testing such as pull-off, thickness and holiday testing.

For non-piggable pipelines, using spray applied solutions were investigated. These are better suited for shorter sections of pipeline (up to 100 m), due to the limited reach of equipment and surface preparation requirements, e.g. abrasive blasting. Longer pipelines would need to be segmented, which is generally unfeasible. Suppliers recommend modifying longer pipelines to make them piggable for effective internal lining.

5.2.4 Typical Preparation & Implementation Steps

The following steps are required to implement in-situ coating of pipelines:

- Desktop review and risk screening: Assess gas composition, operating conditions, presence of valves and bends, isolation capacity and safety constraints. Establish feasibility of process, including confirmation of contractor qualifications and track record on similar applications.
- Detailed pre-survey: Inspect access points, confirm bends and radii, presence of valves, reducers, pipe run distances. Identify pig launch/receive locations or spray rig equipment.
- Coating selection: Choose compatible coating or epoxy system that is compatible with application and gas specification. Trial applications may be considered.
- Identification of cleaning processes and methods, including acceptance standards, as well as assessment of practicality of cleaning methods where cleaning using pigs is not possible.
- Actual execution, including executing cleaning and coating activities, controlling application speed, ensuring safety compliance of equipment.
- Allowance for curing and conditioning after applying coating, including controlling ambient conditions, thermal assist, return to service procedures and checking for trapped volatiles.
- Inspection and testing including pull off tests / holiday testing where access is available, camera of visual inspections
- Documentation and maintenance plan, including documenting as-applied records, inspection photos, establishing maintenance requirements, etc.

5.2.5 Cost Estimates

Budget quotations were obtained for high level cost estimates to support NGT Cost Benefit Analysis (CBA) and comparison with other solutions. Pricing was obtained from Corrocoat, a UK specialist in corrosion protection and in-situ lining. The estimate below covers lining with a Polyglass VE LPLP system for specific pipeline lengths. For additional sections, costs may reduce by 25% due to shared establishment costs. These estimates are indicative only and depend on factors, including the piggability of the pipeline, possible modifications and number of sections to be coated.

See Table 5-1 below for a cost summary.

Description	Diameter (mm)		
	300	500	750
Length (m)	Up to 3000	Up to 5000	Up to 5000
Budget quotation received (GBP)	1,013,063	2,026,100	3,376,870
Additional allowance for Owners Cost (25%), GBP	253,266	506 525	844,218
Total price (GBP)	1,266,329	2,532,625	4,221,088
Price for additional pipeline sections	25% of total price for first section	25% of total price for first section	25% of total price for first section

Table 5-1: High-level Cost for In-situ Internal Lining

6 High Risk Areas and Inspection and Maintenance

6.1 Internal Corrosion Threat Review

Assuming there will be an increased oxygen content, the following locations on the subsurface and above ground pipework have been identified as having an elevated threat of internal corrosion and where additional mitigation may be required.

Location	Operational Considerations contributing to susceptibility	Specific locations with Elevated Susceptibility
Subsurface pipelines from wellheads into storage caverns will be exposed to wet gas conditions with high oxygen level	Pipelines where temperature reduces to below the cavern temperature as wet gas is extracted from storage and water condensate forms.	Upper sections of the vertical subsurface pipelines, pipework and components at the wellhead, as well as surface pipeline sections near the wellheads.
Above ground Pipework	Water drop out	Low points along the buried and above ground pipelines (such as road and other crossings) between wellhead and dehydration facilities

Table 6-1: Susceptible Locations for Corrosion in Service

Additional Areas for consideration:

- Offtakes to adsorption trains (as well as other areas where significant geometry changes occur) or dead legs.
- Equipment and vessels exposed to wet gas as part of the dehydration process before reintroduction back into the NTS. Failure of these systems may allow untreated gas into the network.

6.2 Review of Current Inspection & Maintenance Protocols

Requests for information were issued to UESO members to obtain input on current inspection and maintenance practices in terms of corrosion management.

Feedback was received from a number of members and management strategies varied. Typical mitigation measures that are currently in pace for internal corrosion management are summarised below:

- Ultrasonic Thickness Measurements (UTM) are undertaken at targeted locations, as per the requirements of the PSSR. Locations are selected where moisture is likely to collect, as well as at other higher risk areas. It is not clear if thickness measurements are undertaken on buried pipelines.
- No internal inspections using ILI technology are undertaken on pipelines. Pipelines at a specific facility are reported to be fully piggable, however no pigging has been undertaken due to the perceived risk of the pigging activity, as well as infrastructure being relatively new (11 years) and no significant corrosion being detected during UTM inspections.
- The design of pipelines typically includes a corrosion allowance of 1.6 to 3mm.
- An operator reported that corrosion coupons were installed at certain sites to monitor corrosion rates, however this practice was stopped in 2017 as the benefit of the coupon monitoring was considered to be questionable since UTM measurements are undertaken at 5-year intervals.

- Risk Based Inspection (RBI) techniques are currently used on vessel and pipework by certain operators where critical assessments are undertaken on specific systems. It is not clear if this is applied to buried pipelines, as well as on plant infrastructure.
- Other monitoring techniques, such as wireline techniques for production tubing, internal rotary inspection system (IRIS) and boroscope inspections at heat exchangers, thermography of adsorber vessels and CIPS (close interval potential survey) to monitor the effectiveness of CP systems are reported to be implemented by certain operators.

6.3 Review of Regulatory Requirements

6.3.1 Control of Major Accident Hazard (COMAH) [5]

Considering overall regulatory context and stored volume, one of the principal pieces of legislation is the COMAH Regulations. COMAH sites require regulatory review of a submitted Safety Case, based around detailed safety arrangements and risk assessment (typically quantitative).

The intent of COMAH is to prevent a major accident and reducing risks to "As Low As Reasonably Practicable" may not be considered adequate for an event and the probability ***must be reduced significantly***.

6.3.2 Pressure Systems Safety Regulations (PSSR) [6]

Applicable within each fenceline which focuses on pressure hazards and stored energy. The principal relevant requirement is for a **suitable and sufficient Written Scheme of Examination**.

6.3.3 Pipeline Safety Regulations (PSR) [7]

Applicable outside of fencelines, the Pipelines Safety Regulations require each pipeline to be designed for the chemical processes acting on it, composed of suitable materials, and to be "**maintained in an efficient state**, **in efficient working order and in good repair**".

6.3.4 Other Requirements

- The Gas Act: focusses on commercial aspects;
- The Borehole Regulations: require a health and safety document;
- Gas Safety Management Regulations (GSMR) require a similar Case.

6.3.5 HSE guidance

Salt cavity natural gas storage – consent and operational issues [8] discusses much of the above, and recommends that European standards BS EN 1918 [9] (particularly Parts 3 & 5) are adopted for underground gas storage.

6.3.6 IGM

IGEM guidance, such as SR/25 [10] for zoning is applicable nut note, SR/14 specifically excludes reservoirs.

6.4 Recommended Actions to Support Risk Management

Considering the elevated threat of internal corrosion, it is recommended that the following is considered per site (**if the oxygen content cannot be reduced or where other mitigation (such as internal lining) is deemed unsuitable**):

- Review of the internal corrosion risk assessment.
- Determination of suitable mitigation to monitor or mitigate the threat.
- Perform updates to the Safety Case and Written Scheme, driven by careful review and revision of the relevant risk assessments. This may require more detailed sectionalising than before considering the areas of increased internal corrosion susceptibility.
- Develop a long-term internal corrosion management plan to monitor and manage the threat.

7 Conclusions and Recommendations

7.1 Conclusions

Corrosion Threat Review

An oxygen level in excess of 10 ppm is likely to elevate the threat of internal corrosion. The predicted corrosion threat increases as product temperature decreases. As an example, 0.2 mol % (2000 ppm) at 98 bar pressure is predicted to result in a corrosion rate of 0.3 mm per year at 20° C.

Limiting the oxygen content to 10 ppm is likely to limit the corrosion threat to levels below 0.05 mm per year. However, the presence of carbon dioxide may contribute to elevated corrosion pitting rates.

The most susceptible areas to internal corrosion have been assessed as:

1. **Subsurface pipelines from wellheads into the storage cavern:** Depending on operating conditions, condensation may occur. Wet gas conditions may occur in the form of a water stream along the walls of the pipe in annular mist flow.
2. **Above ground pipework:** Depending on operating conditions, there may be water drop out before dehydration which would result in an elevated corrosion threat at low points.

The corrosion assessment completed in this analysis and the conclusions above are based on a number of scenarios which were deemed representative of the data provided by all underground energy storage operators.

Internal Mitigation Review – Corrosion Inhibitor

This study has reviewed the corrosion inhibition options as possible mitigation. Our evaluation has concluded that options are limited and there is no 'tried and tested' solution in the market to reduce oxygen. Further work would be required to investigate and validate this as a suitable option.

Internal Corrion Mitigation Review – Internal Lining of Pipelines

This study has reviewed internal lining of in situ pipelines as possible mitigation option. Our evaluation has concluded that whilst options are available for short above ground sections, the in-situ lining of subsurface pipelines from the wellhead into the storage facilities was assessed as not currently feasible. Further work would be required to investigate and validate this as a suitable option.

7.2 Recommendations

Corrosion Threat and Risk Management

Considering the elevated threat of internal corrosion at increased oxygen levels, it is recommended that the following is considered per site:

- If not already in place, each of the underground energy storage operators should perform a site-specific corrosion assessment considering their specific pipeline design, historic and future operating envelope to assess the corrosion threat and confirm the most susceptible locations.
- Ensure suitable corrosion mitigation is installed to monitor or mitigate the threat ensuring that the oxygen content is below 10 ppm as a minimum (e.g. the installation of oxygen removal equipment).

- Perform updates to the Safety Case and Written Scheme of examination, driven by careful review and revision of the relevant risk assessments. This may require more detailed sectionalising than before considering the areas of increased internal corrosion susceptibility.
- If not already in place, develop a long-term internal corrosion management plan to monitor the threat.
- Ensure all associated procedures, competency or certification requirements are reviewed and updated if required.

7.3 Suggestions for further Study:

Suitability and Feasibility

If deemed appropriate, perform further evaluation of the mitigation reviewed as part of this study (e.g. corrosion inhibition). This study could review what testing is required to further develop or validate the option, how could these be installed and identification of any uncertainties or risks.

Contingency Planning (site specific)

Investigation of the likely corrosion threat during an operational outage where the selected mitigation has failed (e.g. Oxygen removal equipment or inhibitor injection failure). What are the threats, what levels of oxygen would be likely, where would corrosion occur and how long can the system be operational without mitigation before it becomes a significant issue.

Risk Management (Site Specific)

Perform a corrosion screening assessment across all storage sites to identify the sites likely to have seen the most internal corrosion to date or sites likely to be exposed to elevated corrosion rates in the future considering operating conditions.

In order to baseline the condition, review the feasibility of inspecting the subsurface pipework of the highest risk site with tethered or robotic in-line inspection technology to assess the current condition, confirm how much of the corrosion allowance has been consumed to date and enable routine monitoring during future operation. This would provide reassurance on the other storage sites and help to demonstrate regulatory compliance.

8 References

[1] DNV, “Gas Quality – Impacts and Concentration Limits, Oxygen in the NTS, National Gas Transmission-10514136/1, Rev. 3 and 3a,” 2024.

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[4] D. Tomans, Temperature and Pressure Dependent Solubility of Oxygen in Water: A Thermodynamic Analysis, Hydromrtallurgy 48(3), P. 327-342, May 1998.

[5] Health and Safety Executive (HSE), Contol of Major Accident Hazzard (COMAH) Regulations, 2015.

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[8] Health and Safety Executive (HSE), “Salt Cavity Natural Gas Storage - Consent and Operational Issues, Online Resource,,” [Online]. Available: https://www.hse.gov.uk/gas/supply/saltcavity.htm. [Accessed 2025].

[9] British Standards Institution (BSI), Gas Supply Systems Underground Gas Storage, BS EN 1918, 2016.

[10] The Institution of Gas Engineers and Managers (IGEM), Hazardous Area Classification of Natural Gas Installations, SR/25, Edition 2, 2013.

Appendix A Information Provided by UESO Members on Pipeline Infrastructure

Parameter:	Site Operator 1		Site Operator 2			
	Site 1A -Wellhead - Gas Plant	Site 1A - Subsurface - wells	Site 2A - Surface wellhead - plant	Site 2A - Subsurface - wells	Site 2B - Surface - wellhead - plant	Site 2B - Subsurface -wells
Length	[redacted]					
Diameter (s)						
Material types, grades & wall thickness						
Operating Pressure (Barg)						
Design Pressure (Barg)						
Operating Temperature (deg C)						
Design Temperature (deg C), Min & Max						

	Site Operator 1		Site Operator 2			
Parameter:	**Site 1A -Wellhead - Gas Plant**	**Site 1A - Subsurface - wells**	**Site 2A - Surface wellhead - plant**	**Site 2A - Subsurface - wells**	**Site 2B - Surface - wellhead - plant**	Site 2B - Subsurface -wells
Volume flow rate	[redacted]					
Velocity						
Details of any applicable equipment forming part of system.						
Buried or above ground						

Table A-1: Example of Pipework at Storage Sites (provided by two Operators)

Appendix B Dissolved Oxygen in Water

Dissolved oxygen (ppm)							
Temp (°C)	Pressure (Atm)						
	30	50	60	98	120	250	300
5	10.54	17.34	20.75	27.55	41.15	85.36	102.36
10	9.39	15.44	18.47	24.52	36.64	76.00	91.13
15	8.44	13.89	16.62	22.07	32.96	68.38	82.00
20	7.67	12.62	15.10	20.05	29.95	62.13	74.51
25	7.04	11.58	13.85	18.39	27.47	56.99	68.34
30	6.51	10.71	12.81	17.02	25.42	52.73	63.23
35	6.07	9.99	11.95	15.87	23.71	49.18	58.98
40	5.71	9.39	11.24	14.92	22.29	46.23	55.44
45	5.41	8.89	10.64	14.13	21.10	43.77	52.49
50	5.15	8.48	10.14	13.47	20.12	41.73	50.04
55	4.94	8.13	9.73	12.92	19.30	40.03	48.01

Table B-1: Oxygen in water in contact with gas containing 1 mol% oxygen

Dissolved oxygen (ppm)							
Temp (°C)	Pressure (Atm)						
	30	50	60	98	120	250	300
5	2.11	3.47	4.15	5.51	6.87	8.23	17.07
10	1.88	3.09	3.69	4.90	6.12	7.33	15.20
15	1.69	2.78	3.32	4.41	5.50	6.59	13.68
20	1.53	2.52	3.02	4.01	5.00	5.99	12.43
25	1.41	2.32	2.77	3.68	4.59	5.49	11.40
30	1.30	2.14	2.56	3.40	4.24	5.08	10.55
35	1.21	2.00	2.39	3.17	3.96	4.74	9.84
40	1.14	1.88	2.25	2.98	3.72	4.46	9.25
45	1.08	1.78	2.13	2.83	3.52	4.22	8.75
50	1.03	1.70	2.03	2.69	3.36	4.02	8.35
55	0.99	1.63	1.95	2.58	3.22	3.86	8.01

Table B-2: Oxygen in water in contact with gas containing 0.2 mol% oxygen

Dissolved oxygen (ppm)							
Temp (°C)	Pressure (Atm)						
	30	50	60	98	120	250	300
5	0.011	0.017	0.021	0.028	0.034	0.041	0.085
10	0.009	0.015	0.018	0.025	0.031	0.037	0.076
15	0.008	0.014	0.017	0.022	0.028	0.033	0.068
20	0.008	0.013	0.015	0.020	0.025	0.030	0.062
25	0.007	0.012	0.014	0.018	0.023	0.027	0.057
30	0.007	0.011	0.013	0.017	0.021	0.025	0.053
35	0.006	0.010	0.012	0.016	0.020	0.024	0.049
40	0.006	0.009	0.011	0.015	0.019	0.022	0.046
45	0.005	0.009	0.011	0.014	0.018	0.021	0.044
50	0.005	0.008	0.010	0.013	0.017	0.020	0.042
55	0.005	0.008	0.010	0.013	0.016	0.019	0.040

Table B-3: Oxygen in water in contact with gas containing 0.001 mol% oxygen

Appendix C Corrosion Rate for Different Dissolved Oxygen Levels

Dissolved oxygen (ppm)	Calculated corrosion rate (mm/year)
0	0.00
1	0.08
2	0.15
3	0.23
4	0.3
5	0.38
6	0.46
7	0.53
8	0.61
9	0.68
10	0.76
11	0.83
12	0.91
13	0.98
14	1.06
15	1.14
16	1.21
17	1.29
18	1.36
19	1.44
20	1.51

Table C-1: Oxygen Corrosion Rates in Condensing Water Films

Appendix D Information Provided by UESO Members on Current Inspection and Maintenance

Item	Existing Policie and practices inspections	Operator 1 – Site A	Operator 2 – Site A & B
1	[redacted]		

Item	Existing practices Policie an inspections	Operator 1 – Site A	Operator 2 – Site A & B
2	[redacted]		
3			
4			

Item	Existing Policie practices an inspections	Operator 1 – Site A	Operator 2 – Site A & B
5	[redacted]		
6			
7			
8			

Table D-1: Information provided by UESO members on Current Inspection and Maintenance Practices

Underground Energy Storage Operators
UESO
UESO is a private limited by guarantee company
UESO
13 Thornton Hill
London, SW19 4HU
www.ueso.co.uk
info@ueso.co.uk
Co reg no: 13208286

Introduction
Underground Energy Storage Operators (UESO) represents the owners and operators of underground energy storage facilities in GB. Members have interests in the existing natural gas and future hydrogen storage facilities including both short-medium range salt cavity facilities to longer range onshore and offshore converted gas fields.

UESO members are concerned that as a result of the anticipated increased penetration of biomethane into network that this will result in heightened concentrations of oxygen present in the natural gas injected into gas storage facilities. Increased levels of oxygen in biomethane can lead to corrosion of assets in conditions where wet gas conditions is present, compromising the integrity of storage facilities through accelerated corrosion. Other known consequences of increased levels of oxygen on gas storage include:

- Subsurface and surface sulphur precipitations and resulting corrosion, malfunction of valves and safety devices, plugging of pores and detriments to capacity/availability ;
- Generation of pyrophoric iron sulphide, creating safety issues when exposed to air;
- Blockages caused by oxidation of minerals in formation water; and
- Accelerated degradation through oxidation of glycol in dehydration units.

National Gas NTS is currently seeking HSE consent to increase the GSMR limit for some entry points up to 1mol% O_2. At present the maximum level of oxygen content that such sites may deliver into the NTS is limited by Gas Safety (Management) Regulations 1996 to 0.2mol% O_2. Typically, Storage Connection Agreements entered into by storage operators contain O_2 limits of <10PPM (0.001%).

Natural gas storage facilities in GB
All owners/operators of natural gas storage facilities in GB are members of UESO. Table 1 below provides a summary of the facilities, noting that for the purposes of this project the maximum injection rates are particularly relevant, as any deployed oxygen removal technology at storage connection points would need to sized for the site's maximum injection rate.

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Table 1: Overview of GB natural gas storage facilities

Storage Site	Owner / Operator		Type	Working Gas Volume (declared)	Max. Withdrawal Rate (declared)	Max. Injection Rate (declared)
Aldborough	SSE Hornsea / Equinor		9 Salt Caverns	282 mcm	26 mcm/d	26 mcm/d
Hatfield Moor	Scottish Power		Depleted Gas Field	70 mcm	2 mcm/d	2 mcm/d
Hill Top	Kistos Energy Storage		5 Salt Caverns	59 mcm	13 mcm/d	13 mcm/d
Hole House Farm	Kistos Energy Storage		4 Salt Caverns (suspended)	-	-	-
Holford	Uniper UK		8 Salt Caverns	240 mcm	22 mcm/d	26 mcm/d
Hornsea (Atwick)	SSE Hornsea		8 Salt Caverns	308 mcm	12 mcm/d	3 mcm/d
Humbly Grove	Humbly Grove Energy		Depleted Oil Field	254 mcm	7 mcm/d	8 mcm/d
Rough*	Centrica Energy Storage+		Depleted Gas Field	1,500 mcm	11 mcm/d	9 mcm/d
Stublach	Storengy UK		20 Salt Caverns	400 mcm	30 mcm/d	30 mcm/d

* Rough is currently not operating as a storage facility

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UESO project scope

In order to assess the feasibility of reducing or eliminating oxygen from biomethane and/or introducing corrosion protection technology in wet gas conditions, UESO provided data to project partners.

The data items required to be provided are as stated below:

- Moisture levels
- Operating pressures
- Temperatures
- Flow rates
- Historic corrosion data

Following a Request for Information sent to UESO members and a followup meeting between Frontline Integrrity, National Gas and UESO, the data set out in Table 2 was submitted, covering three salt cavity and one depleted field facilities.

Table 2: UESO data table

Parameter:	Typical Range or set of represenative values to be considered in assessment (note - we have not budgeted to assess multiple sites)	System 1	System 2	System 3	System 4
Length					
Diameter (s)					
Material types, grades & wall thickness					
Operating Pressure (Barg)					
Design Pressure (Barg)					
Operating Temprature (deg C)					
Design Temperature (deg C), Min & Max					
Volume flow rate					
Velocity					
Details of any applicable equipment forming part of system.					
Buried or above ground					

Key observations

The key data items and reported levels in terms of determining the feasibility of applying advanced catalytic and absorption technologies are:

- Critical to oxygen removal considerations are injection rates, pressures and O_2 concentrations;
-

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- Pipeline lengths exposed to ‘wet gas’ on studied sites ranged from 1km to 20km;
- Operational pressures in storage varied from 15 barg to 286 barg; and
- Pressures at NTS typically operating between 50-65 barg (for O_2 removal).

The data presented represents a sample of four storage facilities. These facilities exhibit significant variation in pipeline lengths and operational pressures, which is representative of the broader storage community.

GB storage facilities have been constructed over an extended period, utilizing the technology and resources available at each stage of development. Beyond the challenges posed by extensive pipeline lengths containing 'wet gas' and variable operational pressures, the diverse characteristics of the storage fleet preclude a standard "one-size-fits-all" solution.

Challenges for oxygen removal at storage connection points:
Based on discussion with project partners and following initial reviews of the date presented the following challenges to introducing existing oxygen removal technologies are as follows:

- High Pressures required for gas flows in to and out of storage;
- High Flow rates relating to volumes and velocities;
- Challenges for corrosion inhibition at storage sites in terms of gaining access to the impacted infrastructure and the overall length of effected pipelines etc;
- Range of pressure (including frequent cycling);
- Length of affected piping systems, particular involving wet gas; and
- Chemical interaction with hydrate prevention dosing.

Conclusions
Based on the data presented and initial discussions with project partners, it was concluded that the use of oxygen removal technologies and the application of corrosion inhibitors was not practical from either a technical or economical perspective. In order to limit the incidence of increased oxygen concentrations in biomethane, and thereby reducing the corrosive impacts on storage facilities, resources would be more effectively deployed at the point of entry of biomethane into the gas network i.e. the biomethane production facilities.