Biomethane production using an integrated anaerobic digestion, gasification and CO_2 biomethanation process in a real waste water treatment plant: A techno-economic assessment

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Abstract. The biomethanation of CO_2 from anaerobic digestion within the power to gas concept has recently emerged as a promising technology to upgrade biogas, to decarbonise the domestic and industrial heat sector, provide long term energy storage and deliver grid balancing services. In addition, the utilisation of the digestate, through a process such as gasification, offers a circular economy approach and has the potential to enhance the deployment of power to gas systems. To this direction, the study focuses on exploring the techno-economic feasibility of coupling biomethanation with digestate gasification for the wastewater industry. The study constitutes the first endeavour to assess the viability of such an integrated energy system. Four different scenarios have been designed and assessed. The energy efficiency of the concepts lies between 26.5% and 35.5% while the minimum selling price (MSP) of biomethane is in the range of 135-183 £/MWh. The implementation of appropriate policy mechanisms and the inclusion of by-products revenues reduces the MSPs by approximately 32%-42%. The conduction of a typical sensitivity analysis has identified the electricity price as the prime cost driver and this is followed by the cost of the electrolyser or the gasification plant depending on the scenario. Finally, a 2030 analysis, that incorporates projected techno-economic advances, has been carried out and revealed that under certain circumstances profits can be generated.

Keywords: biomethanation, power to gas, biogas upgrading, CO_2 utilisation, techno-economics circular economy

1. Introduction

The Paris Agreement commits signatories to holding the increase in the global average temperature to well below 2°C degrees above pre-industrial levels, and to pursue efforts to limit the temperature increase to 1.5°C degrees above pre-industrial levels [1]. In view of this, resultant environmental concerns and rising global population in conjunction with the augmenting global demand for energy, chemicals and materials have facilitated research efforts to advance low carbon technologies contingent on renewable sources to encounter such global targets [2]. To this direction, attention has been drawn by the research community to effectively integrate renewable technologies and promote alternative raw materials that can potentially substitute fossil feedstocks [3]. Furthermore, a distinct advantage of modern integrated energy systems is that they possess the potential to foster the acceleration of the transition from linear to circular economies.

The deployment of anaerobic digestion (AD) to produce biomethane has been recognised as an efficient and promising route over fossil fuels that can mitigate environmental impacts [4] and give rise to a product that has multiple applications (*e.g.* electricity, heating, transport) [5]. Nevertheless, the generation of high purity biomethane via the AD process can be only achieved by incorporating an additional step that serves to upgrade the original product of AD, *i.e.* biogas, to biomethane. Typically, biogas comprises 50-70% methane and 30-50% carbon dioxide and is commonly utilised for local combined heat and power (CHP) applications [6]. However, this composition does not allow direct injection to the prevailing natural gas (NG) networks and thereby advantages related to the efficient storage of biomass derived energy and on-demand energy utilisation as well as the stabilisation of an energy supply system that is based on renewable intermittent sources cannot be realised [7]. Furthermore, the utilisation of biomethane has the potential to offset the reliance on NG imports and therefore enhances energy procurement security. It has been reported that a great potential in primary energy savings for national economies exists if biogas from large scale plants is to be upgraded to biomethane for substituting fossil fuels in place of utilising it for generating renewable electricity [8]. In several countries biogas plants are anticipated to move from electricity generation to biomethane manufacture, provided that sufficient policy support is available [7].

Among the different options to upgrade biogas, such as chemical/physical absorption, membrane and cryogenic technologies [9][10], the Power to Gas (P2G) concept has received increased interest in the recent years [11]. The idea behind the P2G technology is that electricity can be utilised to hydrolyse water and the produced hydrogen will react with the CO_2 in the biogas to form methane though the biological Sabatier reaction, Eq.(1).

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O, \;\; \Delta H^0 = -165 \; kJ/mol \quad (1)$$

The P2G can provide long term storage of excess renewable energy in existing natural gas infrastructures offering grid balancing services and increasing the biomethane yield [12]. Even if a P2G process that targets methane production over hydrogen achieves lower thermodynamic efficiency, the existing gas grids are designed for methane distribution and the direct injection of hydrogen can be limited and thus methane may be a more suitable and effective option [13].

Two approaches exist to produce methane via the Sabatier reaction, *i.e.* chemical and biological synthesis [14]. The latter operates at much lower process conditions (*e.g.* temperature and pressure) and can treat biogas of less strict quality [15]. In addition, the reactor design and conditions can be easier adjusted to the requirements of the biomethanation and it is appropriate for small-scale applications such as the utilisation of CO_2 from AD [16]. Even if the biological methanation is less mature than the chemical methanation, there are a couple pilot and demo plants that presently examine the performance of biomethanation due to the above mentioned advantages. The MicrobEnergy plant in Schwandorf and the BioCat project in Copenhagen are examples of successful implementation of power to gas systems via biomethanation [16].

Although AD is an efficient way to convert organic matter into energy, it deals only partly with the issue of material and energy recovery. This is because a substantial portion of the organic matter is not biodegradable and remains in the residual material of the AD, known as digestate [17]. The common way to valorise the digestate is by using it as a soil amendment and/or fertilizer provided that it meets the constraints imposed by the European Nitrate Directive (91/676/EEC) [18]. In addition, the increase in the number of AD facilities in certain regions that exhibit extensive livestock farming, is projected to result in a surplus of digestate supply [19]. It is subsequently expected that these agricultural regions will not be able to treat these augmented amounts of digestate and hence transport to other areas, that display nutrients shortfall, will be required; a practice that will increase the operating expenditures of AD plants [20]. As a result, extensive uptake of regional scale AD is constrained because of financial obstacles and the guarantee of the harmless disposal of the digestate. To address these issues, the scientific community has recently drawn attention to other digestate valorisation options including gasification [20] [21] and pyrolysis [22][23]. These processes are capable of converting non-biodegradable materials (*e.g.* digestate), thermochemically at elevated temperatures, into valuable products including syngas, bio-oil, fuels and chemicals [24]. This approach can facilitate the expansion of AD based infrastructures and give a boost to infrastructures that promote circular economies.

The conduction of a literature review revealed that previous studies have examined the feasibility of power to methane production, either via chemical [25][26] or biological synthesis [27][28], but none

of them consider the valorisation of digestate. In addition, scarce data exists regarding the economic appraisal of integrated AD-thermochemical routes systems; Salman *et al*. [29] have explored the possibility of coupling AD with pyrolysis but only from an energetic/exergetic point of view while Li *et al.* [30] tested the economic performance of a system, which combines AD and gasification, that process two types of feedstock, namely household waste and wood pellets, and thereby a linear rather than a circular energy utilisation system was considered. In addition, the incorporation of P2G technologies was out of the scope of both these studies [29][30].

In view of the above, the study attempts to appraise the technoeconomic performance of a P2G system that closes the energy and material loops of an AD plant and produces high purity methane. In more detail, biogas produced via anaerobic digestion of sewage sludge in a real waste water plant (WWTP) is upgraded through biomethanation to methane. Depending on the investigated scenario, the whole or part of the hydrogen required for the latter process is generated through water electrolysis while the remaining demand is covered by the hydrogen produced by means of digestate gasification. The research is the first endeavour to assess the feasibility of coupling AD with gasification within a P2G concept. The findings of the study can aid to identify novel methods that can enhance the profitability of waste water plants in a sustainable manner. In turn the results of the study can also inform policy makers on the new regulatory frameworks and incentives required to enable the deployment of power to gas technologies.

2. Research outline

2.1. System boundaries and definition of scenarios

The study assesses the techno-economic performance of upgrading biogas produced in an existing WWTP via biomethanation. The aim of the work is to evaluate retrofit opportunities for biomethane production in a typical WWTP. The WWTP treats sewage sludge by means of anaerobic digestion. It is located in the UK and operated by United Utilities. The plant includes eight digesters of identical size. The study considers the upgrade of the biogas derived from two of them.

The system boundaries for the engineering design include a pair of twin digesters, a biomethanation reactor that upgrades biogas to biomethane through the biological Sabatier reaction, a proton exchange membrane (PEM) water electrolysis unit, a digestate dewatering unit and the digestate utilisation unit which is an entrained flow (EF) gasification plant configured either to produce hydrogen or generate electricity by means of a typical integrated gasification combined cycle (IGCC) system. The sizing of the power to gas components were based on the size of the AD plant. Therefore, it was necessary for the quantification of the material streams to expand the boundaries of the study to include existing units. On the other hand, for the economic evaluation, since all the

cases investigated herein are retrofit projects, the boundaries include only the electrolysis unit, the biomethanation reactor and the digestate gasification and excludes the digesters, the CHP unit and the dewatering unit because these already exist as an integral part of the WWTP. Nevertheless, the current utilisation of the biogas is to raise electricity of 2.4 MW that covers internal needs of the plant and thereby this amount of electricity has to be purchased now in the retrofit design. Thereby, an opportunity cost arises that has been included in the calculation of the OPEX.

Biomethanation can be applied *in-situ*, when it takes place in the digester or *ex-situ* when it occurs in a separate reactor. The former approach can only accomplish a final methane composition of up to 90% [31] while the *ex-situ* system can realise a high purity methane stream, *i.e.* ~98% [32], that can meet the stringent conditions of the NG grid network. Furthermore the *ex-situ* technology is of a higher TRL (technology readiness level) system with demo plants being already in operation [16] while the in-situ is only proven at lab scale. On considering the above discussion, it was decided to investigate the viability of an *ex-situ* biomethanation plant.

For the selection of the digestate gasifier there are many available options, such as fixed bed, fluidised bed, entrained flow and plasma gasifiers, and each has its own advantages and disadvantages [33]. Herein, an entrained flow gasifier has been selected. The gasification oxidising agent is a mixture of steam and oxygen; oxygen is supplied from the electrolyser. The advantages of the EF over other gasification technologies include low tar and methane formation, high carbon conversion, low residence time and reduced equipment size [34][35]. Nevertheless, the main drawback of the EF is the necessity to pulverise the fuel to extremely low particle size making the feedstock pretreatment more energy and cost intensive compared to other gasification technologies [36]. It is not the intention of the study to advocate a specific type of gasifier as the most suitable for digestate gasification but only to carry out preliminary mass and energy balances for digestate gasification. It is, therefore, recommended that future research could focus on exploring the most suitable technology for digestate gasification.

Figure 1 is a visual representation of the scenarios examined in the study by means of simplified block flow diagrams (BFD). In all scenarios the size of the digesters remains constant while based on the particularities of each case the size of the biomethanation reactor and the PEM unit vary accordingly. Four cases in total were investigated and these are defined as follows:

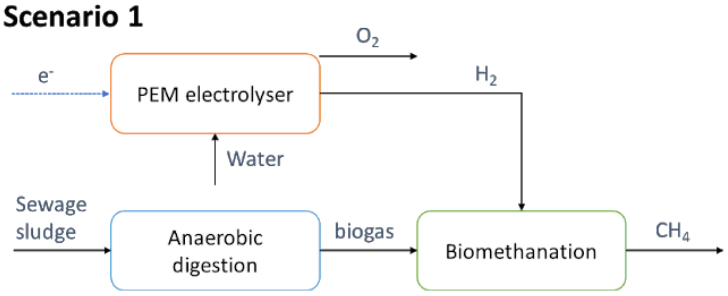
Scenario 1 (BioMeth): This scenario is the simplest one and serves as the basis for further comparisons. It does not consider the digestate utilisation and thereby the whole amount of H_2 is provided by the electrolyser. The digestate is dewatered to a sludge cake 25% dry solids and then recycled to farm land.

Scenario 2 (BioMeth+EF-H_2): Here the digestate is gasified and the produced syngas is treated appropriately in a series of water gas shift reformers to give rise to H_2. The H_2 is supplied to the biomethanation reactor which means that the electrolyser size is reduced (compared to Scenario 1) and less electricity is needed.

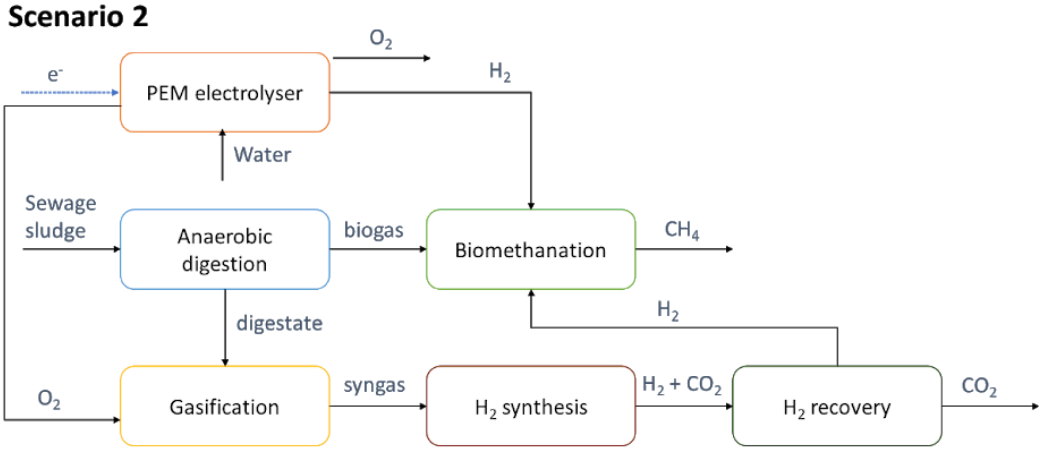
Scenario 3 (*BioMeth+EF-CH_4)*: The incorporated technologies are identical with the Scenario 2 but herein instead of supplying H_2 to the biomethanation reactor a mixture of H_2 and CO_2 (ratio 4:1) is provided for enhanced methane production. Therefore, the size of the biomethanation reactor increases (compared to Scenarios 1 and 2) while the size of the electrolyser remains unchanged (the same as in Scenario 1).

Scenario 4 (*BioMeth+IGCC)*: The last scenario assumes an IGCC plant with the aim of generating electricity that will partially cover the electricity demand of the electrolyser. In this case, syngas is combusted in a gas turbine while the waste heat is utilised to raise steam that will drive steam turbines. The electrolyser and the biomethanation reactor are of identical size as in the Scenario 1.

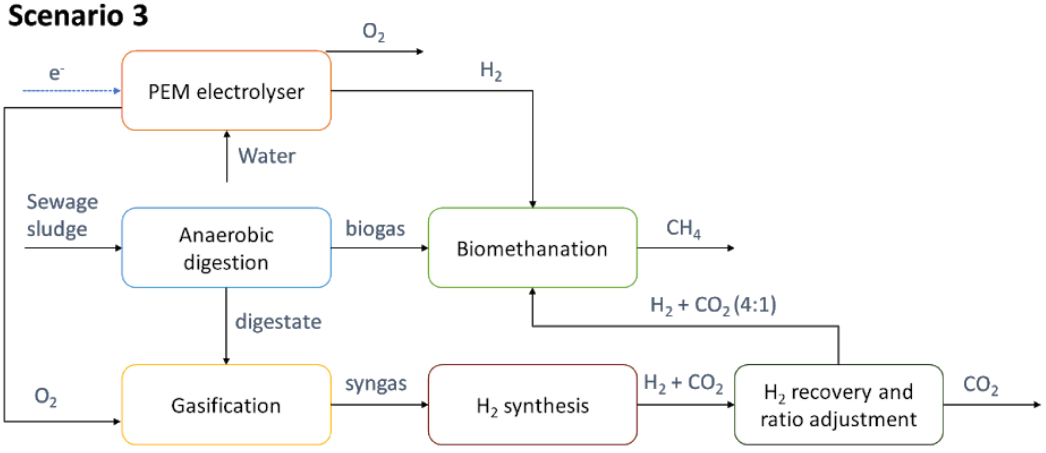
Scenario 1



Scenario 2



Scenario 3



Scenario 4

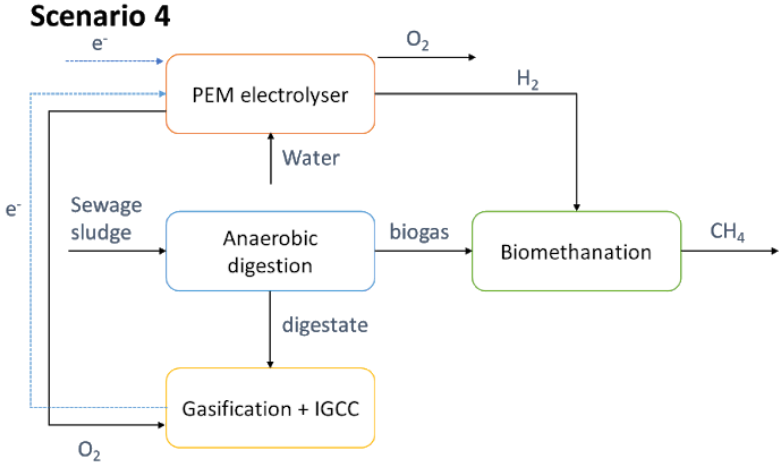


Figure 1. Block flow diagrams of the investigated scenarios

Overall, the aim of the present research is to carry out comprehensive techno-economic assessments and comparisons of novel energy systems that endorse circular economies and subsequently to detect key cost drivers and conditions under which they can achieve commercialisation.

2.2 Basis for process modelling

A conceptual design of the proposed configurations was carried out to quantify the mass and energy flows. The AD unit was designed based on data retrieved from a real WWTP and balance calculations were conducted in Excel. The PEM electrolyser was modelled in Simulink/Matlab environment with the aim of estimating the cell voltage and the stack energy efficiency. The digestate gasification and combustion plants have been modelled in Aspen Plus by considering three models in total to define the thermodynamic properties: Redlich-Kwong-Soave with Huron-Vidal mixing rules for high pressure streams (>10 bar), NRTL-RK for low pressure streams (<10 bar) and classic steam tables for the CHP units [37][38]. Lastly, **Table 1** summarises the technical specifications of typical equipment [39][40].

Table 1. Specifications for common equipment utilised in the present design [39][40].

Equipment	Parameter	Value
Gas turbine	Isentropic efficiency	90%
HP steam turbine	Isentropic efficiency	92%
MP steam turbine	Isentropic efficiency	94%
LP steam turbine	Isentropic efficiency	98%
Turbines	Mechanical efficiency	98%
Pump	Efficiency	70%
Heat exchanger	Temperature difference	7°C
Cooling water	Temperature range	15-25°C

2.3 Basis for the economic evaluation

As depicted in **Table 2** [35][41][42], a typical bottom-up methodology is used to appraise the capital expenditures (CAPEX) where all cost components are articulated as a fraction of the purchased equipment costs (PEC). In addition, assumptions for estimating operating expenses (OPEX), which consist of fixed and variable costs, are tabulated in **Appendix A Table A.1** [43][44][45] **and Table A.2** [43][46][47][48][49][50]**,** respectively. It should be noted that the typical treatment of the digestate is transportation to farmland for recycling; the cost for this practise is 10 £/t. Also, the residual water after the dewatering unit returns to the WWTP; the cost associated with its management is not included in the study as it is already paid in the business as usual model of the WWTP. The scenarios that gasify the digestate avoid this cost and in the cash flow analysis this appears as revenue. As mentioned before, all scenarios include an opportunity loss (OppLoss) associated with the loss of the biogas derived electricity (existing utilisation of biogas).

In addition, the overall amount of the necessary labour hours has been calculated according to best practice functions related to highly automated fluid processing plants based on the work conducted by Peters *et al.* [44] as follows**,** Eq.(2)**:**

$$h_{labor}\left[\frac{h}{year}\right] = 2.13 \times plant\ capacity\ \left[\frac{kg_{fuel\_output}}{h}\right]^{0.242} \times n_{process\_steps} \times \frac{h_{plant\_operation}}{24} \qquad (2)$$

The plant capacity is expressed in terms of product output, *i.e.* CH_4 in the present study, the number of process steps, $n_{process\_steps}$, refers to segments (one or more unit operations) within the production line wherein significant chemical composition and/or thermodynamic change occurs and $h_{plant\_operation}$ is the annual operating time of the plant. The hourly labour compensation cost was taken as 15 £/h [51].

Table 2. CAPEX estimation methodology [35][41][42]**.**

Cost component	**Factor**
Direct costs (DC)	
Purchased equipment cost (PEC)	100% PEC
Purchased equipment installation	39% PEC
Instrumentation and controls	26% PEC
Piping	31% PEC
Electrical systems	10% PEC
Indirect costs (IC)	21.9% DC
Fixed Capital Investment (FCI)	**DC+IC**
Start-up costs (SUC)	5% FCI
Interest during construction (IDC)	Computed

CAPEX	**FCI + SUC + IDC**
*Working Capital (WC)	5% FCI

**Working capital is applied in the year before operation and recovered at the end of the plant life and hence it is not depreciated*

The scaling method, as defined in Eq.(3), was used to estimate the purchased equipment cost. This equation permits us to obtain a cost for an equipment item of a different size when the cost for a given size is known. Where necessary the costs were updated to GBP2017 by using the Chemical Engineering Plant Cost Indices (CEPCI). Also, the reference equipment cost data was retrieved from the literature and is presented in **Appendix A Table A.3** [27][52][53][54][55][56].

$$C = C_0\left(\frac{S}{S_0}\right)^f \tag{3}$$

Where

$C = actual\ cost$

$C_0 = base\ cost$

$S = actual\ capacity$

$S_0 = base\ capacity$

$f = scaling\ factor$

As depicted in **Table 3**, several assumptions have been utilised for the conduction of a typical discounted cash flow analysis (DCFA). The purpose of DCFA is the estimation of the break-even biomethane price (NPV=0 or IRR=discount rate) also known as minimum selling price (MSP). The basis of the DCFA model is outlined in Eq.(4) and Eq.(5) [57].

$$\sum_{n=1}^{20} \frac{CF}{(1+IRR)^n} = 0 \tag{4}$$

$$CF_n = P_n(1-t) + D_n t \tag{5}$$

Where*, CF* values are the after tax cash flow for each year, *n* is the number of years, *P* are gross profits, *t* is the tax rate and *D* the depreciation. In addition, the levelised cost of energy (LCOE), in terms of methane HHV and in current GBP, was also estimated as presented in Eq.(6) [58].

$$LCOE\left(\frac{£}{MWh}\right) = \frac{\sum_{n=0}^{20} \frac{costs\ in\ year\ n}{(1+discount\ rate)^n}}{\sum_{n=1}^{20} \frac{MWh\ of\ methane\ produced\ in\ year\ n}{(1+discount\ rate)^n}} \tag{6}$$

Table 3. Economic parameters and assumed values.

Parameter	**Value**

Location	United Kingdom (UK)
Currency	GBP
Base year	2017
Project lifetime (y)	20
Construction period (y)	3
Plant availability (h/y)	8,000
Tax rate (%)	30
Discount rate (%)	10
Depreciation	straight-line
Depreciation period (y)	10
Salvage value (£)	0

3. Process design and modelling

3.1 Anaerobic digestion and biomethanation

As mentioned above, the AD plant was modelled utilising data retrieved from a real WWTP. The work has been done in collaboration with United Utilities who supplied the data presented in **Table 4**. The total sludge flowrate, processed by a pair of twin digesters, is 593.75 m^3/d. The volume of each digester is 8,000 m^3 and they operate at mesophilic conditions of 39°C and atmospheric pressure. The heating demand for the AD bioreactors is taken as 10% of the energy content of the biogas [59]. We have calculated the heating duty of the digesters in order to demonstrate potential heat integration opportunities with the upgrading section as waste heat can be generated in the

proposed system. The sources of the heat exist in the biomethanation reactor as the Sabatier reaction is exothermic and the PEM electrolyser.

The total dry solids (TDS) content is 10.5%w/v whilst the volatile solids (VS) account for 75% of the TDS. The achieved biogas yield is 428 m^3/TDS with a dry composition of 61% for methane and 39% for CO_2. The mass balances developed in the present study did not consider the formation of H_2S as it is present at very low concentrations. Nevertheless, the removal of H_2S is crucial to meet the natural gas grid requirements [60] and therefore even if the technical design does not appreciate H_2S production, the cost evaluation considers a ZnO adsorption unit [53] to remove any H_2S traces. Furthermore, the produced digestate undergoes mechanical dewatering via centrifugation to increase the solids fraction to 25%.

Table 4. Technical data for modelling AD. Data from a real WWTP.

Parameter	Value	Unit
Reactor volume	8,000	m^3
Number of reactors	2	-
Sludge flowrate	593.75	m^3/d
TDS content	10.5	%
VS content	75	% (of TDS)
Biogas yield	428	m^3/TDS
Biogas composition (dry basis)	61/39	% CH_4/CO_2
Digestate solid fraction (after dewatering)	25	%

Thereafter, the biogas is fed to the biomethanation reactor along with H_2 from the electrolysis unit and depending on the scenario from the gasification plant. In any case, the feed ratio of H_2 to CO_2 is always 4:1, as Eq.(1) suggests (Sabatier reaction). The bioreactor operates at mesophilic conditions

of 60°C and pressure of 5 bar [61]. The pressure operating window for biomethanation lies between 1 and 9 bar [61]; we have selected 5 bar as this is the delivery pressure of biomethane to the grid and as such no additional costs for compression are required. A high conversion of CO_2 to methane of approximately 98.6% has been reported in demonstration plants [12] and this value was adopted here. Furthermore, the electrical demand of the biomethanation reactor was taken as 0.45 kWh/m^3 of methane produced [61]. The reactor is a jacketed exothermic CSTR that utilises circulated water to ensure isothermal conditions.

3.2 PEM electrolysis

PEM water electrolysis technology is often demonstrated in the literature as a possibly very effective alternative to the more conventional alkaline water electrolysis. The main advantages are flexibility in operation and higher energy efficiency [62]. The overall electrolysis reaction is the sum of the two electrochemical half reactions, that occur at the electrodes in an acidic environment according to the succeeding reactions [11]:

$$H_2O\,(l) \rightarrow \tfrac{1}{2}O_2\,(g) + 2H^+ + 2e^- \quad (7)$$

$$2H^+ + 2e^- \rightarrow H_2\,(g) \quad (8)$$

Eq.(7) signifies the anode half reaction and Eq. (8) the cathode half reaction. At the anode (positively charged electrode) water is oxidized, the electrons pass through the external electrical circuit and oxygen evolves as gas (oxygen evolution reaction, OER). Protons migrate through the acidic electrolyte from the anode to the cathode (negatively charged electrode) where they are reduced by the electrons from the external electrical circuit to hydrogen [63].

The voltage of the electrolyser, V, is the sum of the open-circuit voltage, V_{oc}, and three overpotentials namely activation, V_{act}, Ohmic, V_{ohm} and concentration, V_{conc} [64]. The latter can be neglected for current densities up to 3 A/cm^2 [64] and this assumption was considered here. Therefore,

$$V = V_{OC} + V_{act} + V_{ohm} \quad (9)$$

The V_{oc} is given as follows [65]:

$$V_{OC} = E^0 + \frac{RT}{zF} ln\left(\frac{p_{H_2}}{P_{cat}}\sqrt{\frac{p_{O_2}}{P_{an}}}\right) \quad (10)$$

Where E^0 reversible cell voltage and is expressed as [65]:

$$E^0 = \frac{\Delta G_R^0}{zF} = 1.229 \quad (11)$$

The partial pressures of H_2, p_{H_2}, and O_2, p_{O_2}, are estimated using Eq.(12) and Eq.(13) [65] as a function of the saturation pressure of water, $P_{sat,water}$. The saturation pressure is temperature dependent and is calculated using Eq. (14) [64]. Also, it was assumed here that the pressure in the anode and cathode are equal [65].

$$p_{H_2} = P_{cat} - P_{sat,water}\ [Pa] \quad (12)$$

$$p_{O_2} = P_{an} - P_{sat,water}\ [Pa] \quad (13)$$

$$P_{sat,water}(Pa) = -2846.4 + 411.24T - 10.554T^2 + 0.16636T^3\ (T\ in\ °C) \quad (14)$$

The activation overpotentials in the anode and the cathode are defined as [66]:

$$V_{act,an} = \frac{RT}{o_{an}zF} ln\left(\frac{i_{an}}{i_{0,an}}\right) \quad (15)$$

$$V_{act,cat} = \frac{RT}{o_{cat}zF} ln\left(\frac{i_{cat}}{i_{0,cat}}\right) \quad (16)$$

Ohmic losses, V_{ohm}, are caused due to ionic and electronic resistances. However, the ionic losses dominate the Vohm and thereby electronic losses were neglected from the current model. As depicted in Eq.(18), the Ohmic resistance is a function of the thickness, t_{me}, and the conductivity, σ_{me}, of the membrane [64]:

$$V_{ohm} = iR_{ohm,i} \quad (17)$$

$$R_{ohm,i} = t_{me}/\sigma_{me} \quad (18)$$

The conductivity, Eq.(19) [64], depends on the temperature and the mean of the water uptake of the membrane, λ_m, in the anode and the cathode. Eq. (21) [64] shows that the water uptake in both sides is a function of the water vapour activity, a.

$$\sigma_{me} = (0.00514\lambda_m - 0.00326)\, exp[1268\left(\frac{1}{303} - \frac{1}{T}\right)] \quad (19)$$

$$\lambda_m = (\lambda_{an} + \lambda_{cat})/2 \quad (20)$$

$$\lambda_i = 0.043 + 17.81a - 39.85a^2 + 36a^3,\ (i = an, cat) \quad (21)$$

$$a = \frac{P_i}{P_{sat}},\ \ (i = an, cat) \quad (22)$$

Eq. (23) [65] presents the electrolyser second law efficiency. This accepts as input the electric work, *W* (in J/mol), supplied to the electrolyser and as output the Gibbs free energy change in standard

condition of the electrochemical reaction to obtain H_2. As shown in Eq.(24), the work is expressed as a function of the cell voltage and the faradaic efficiency [65], η_F . The faradaic efficiency is typically very close to unity and thereby a value of 99% was selected in the present study [65].

$$\eta_{\Delta G} = \frac{\Delta G_R^0}{W} \tag{23}$$

$$W = 2F\frac{V}{\eta_F} \tag{24}$$

The above mathematical expressions constitute the mass and energy balances of the PEM unit and these were applied in Matlab/Simulink environment. The necessary data to develop and execute the PEM model is presented in **Table 5** and it was retrieved from the literature [64][65][66].

Table 5. Fixed model parameters for the PEM electrolyser model [64] [65] [66]

Parameter	Symbol	Value	Unit
Gibbs free energy change	ΔG_R	237.2	kJ/mol
Number of electrons transferred in the cell	z	2	-
Faraday constant	F	96,485	C/mol
Anode pressure	P_{an}	5	bar
Cathode pressure	P_{cat}	5	bar
Cell temperature	T	353	K
Anode charger transfer coefficient	o_{an}	0.5	-
Cathode charger transfer coefficient	o_{cat}	0.5	-
Anode current density	i_{an}	2	A/cm^2
Cathode current density	i_{cat}	2	A/cm^2
Anode exchange current density	$i_{0,an}$	2×10^{-7}	A/cm^2
Cathode exchange current density	$i_{0,cat}$	2×10^{-7}	A/cm^2
Membrane thickness	t_{me}	175×10^{-4}	cm
Faradaic efficiency	η_F	99	%

3.3 Gasification plant

Mass and energy balances for the digestate gasification plant were established in Aspen plus. The composition of the dewatered digestate is depicted in **Table 6** [67]. A user defined non-conventional solid was selected to denote the digestate. To develop this module two Aspen models were apportioned: one for density (DCOALIGT) and one for enthalpy (HCOALGEN) that necessitate the awareness of proximate analysis and ultimate analysis of the feedstock which have been obtained from [68].

Table 6. Digestate composition [67]

Proximate analysis (%)	
Parameters	**Mass fraction**
Ash	43.15
Volatile matter	51.75
Fixed Carbon	5.1
Ultimate analysis (%)	
Element	**Dry Weight**
C	30.88
H	4.36
O	15.61
S	1.24
N	4.76
Ash	43.15

As depicted in **Figure 2**, prior to gasification, the digestate has to be dried in order to reduce the moisture content to 5% and subsequently its particle size is decreased to 1mm via grinding. A rotary direct steam drying unit is considered herein with a steam to evaporated moisture ratio of 9 [69]. Steam enters the dryer at a temperature of 200°C, exits the unit at 120°C and is reheated in the syngas cooling section [69]. Moreover, a gyratory crusher, with a specific electrical energy consumption of 50 kWh/t [70] of digestate, is employed to achieve the desired particle size ensuring efficient heat transfer in the gasifier.

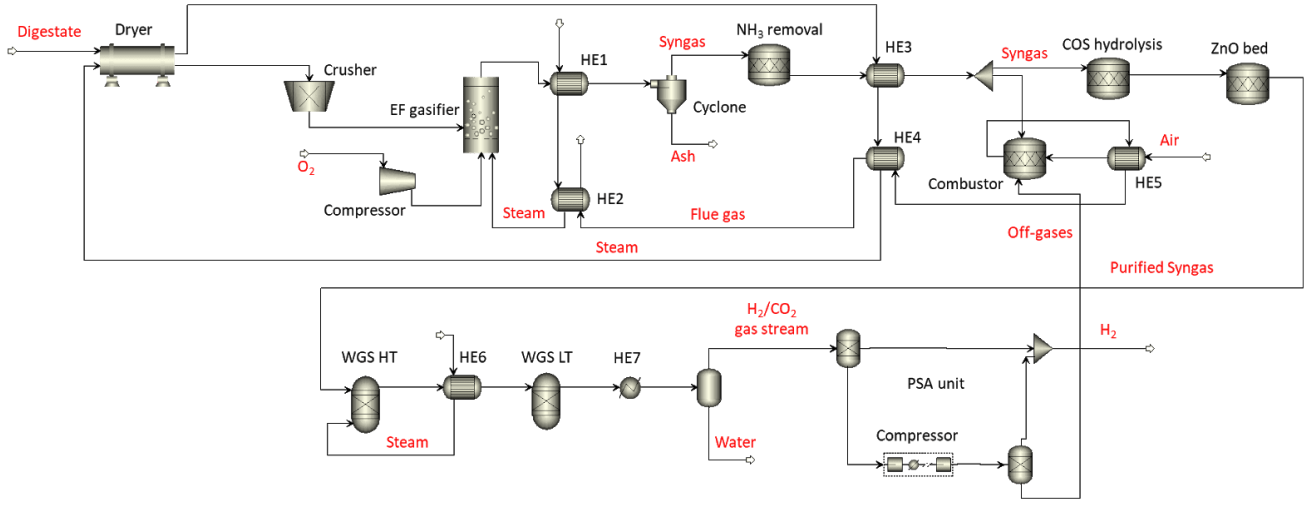


Figure 2. Simplified process flow diagram for the gasification plant aiming to H_2 production in Scenario 2

The EF gasifier was modelled considering two blocks: i) an RYIELD reactor which serves to decompose the digestate to its elements based on the ultimate analysis, and ii) an RGIBBS reactor that predicts the syngas composition assuming chemical equilibrium according to the Gibbs free energy minimization technique. It has been reported in the past [71] that the calculation of the chemical equilibrium conditions for EF gasifiers can give good results. The gasifier operates at 1300°C and 30 bar [34] [72] while the oxygen to biomass ratio (dry basis) is 0.2 and the steam to biomass ratio is 0.19 (dry basis); these ratios were manipulated in such a manner so as the overall heat duty of the gasifier to be zero or equally the heat duty of the RYIELD reactor to match the heat duty of the RGIBBS reactor (detailed mass balances for the gasifier can be found in the Supplementary material). The required oxygen is supplied from the electrolysis unit while steam is raised by recovering heat from the syngas and the flue gas from the combustor unit. Thereafter syngas has to be conditioned to remove impurities such NH_3, COS and H_2S. A high temperature (=800°C) thermal catalytic (nickel based) unit [73] decomposes ammonia to nitrogen and hydrogen while a COS hydrolysis unit converts COS to H_2S and CO_2 [74]. The final purification step assumes a zinc oxide adsorption unit for H_2S removal based on the $ZnO+H_2S \rightarrow ZnS+H_2O$ desulphurisation reaction [75]. The three purification steps were modelled as conversion reactors, RSTOICH, assuming 100% removal of the impurities. A small portion of the syngas (~10%) is purged and sent to a combustor unit to generate heat mainly to cover the heat demand of the drying process.

Thereafter, the treated syngas enters the hydrogen synthesis area whereby CO and high pressure steam are converted to H_2 and CO_2 according to the water gas shift (WGS) reaction, Eq.(25).

$$CO + H_2O \rightarrow H_2 + CO_2, \quad \Delta H^0 = -41\ kJ/mol \tag{25}$$

The WGS is exothermic and as such it is favoured by low temperatures. In order to achieve high H_2 yields, two reactors in series were considered with intermediate cooling. The cooler serves to generate steam which is subsequently supplied to the first reactor. The latter operates in a high temperature (HT) regime of 400°C and a pressure of 30 bar while the second one, in the low temperature (LT) regime, at 200°C and at the same pressure. The REQUIL Aspen plus reactor module that assumes equilibrium conditions was used to simulate the WGS reactors; both reactors operate adiabatically. The gas hourly space velocity of the WGS reactors was taken as 1,000 h^{-1} [69]. The produced gas stream, consisting of H_2 and CO_2, undergoes cooling with the aim of condensing the water content and afterwards it is sent to a typical pressure swing adsorption unit (PSA) to recover H_2. The PSA unit was modelled as a typical separator whereby a H_2 recovery of 90% is achieved. For the Scenario 2, the hydrogen rich stream (>99.9% purity) is sent to the biomethanation reactor while the off-gases are sent to the combustion unit. For the Scenario 3, the purified hydrogen stream is

mixed with part of the PSA off-gases stream (~43%) with the aim of attaining a H_2:CO_2 ratio of 4 and then the resultant gas stream is supplied to the biomethanation reactor. The PSA off-gases stream is a CO_2 rich stream with some hydrogen.

Specifically for the Scenario 4, as depicted in **Figure 3**, the syngas enters a power generation unit comprising a gas turbine and a heat recovery steam generation (HRSG) system for enhanced electricity generation. The syngas is combusted with excess air in a typical gas turbine while heat from the exhaust flue gas is recovered in a series of three heat exchangers namely superheater, evaporator and economiser to raise steam and drive a configuration of high, intermediate and low pressure steam turbines. In this case, flue gas exiting the power island assists in raising steam for the dryer while the oxidising agent of the gasifier is solely oxygen.

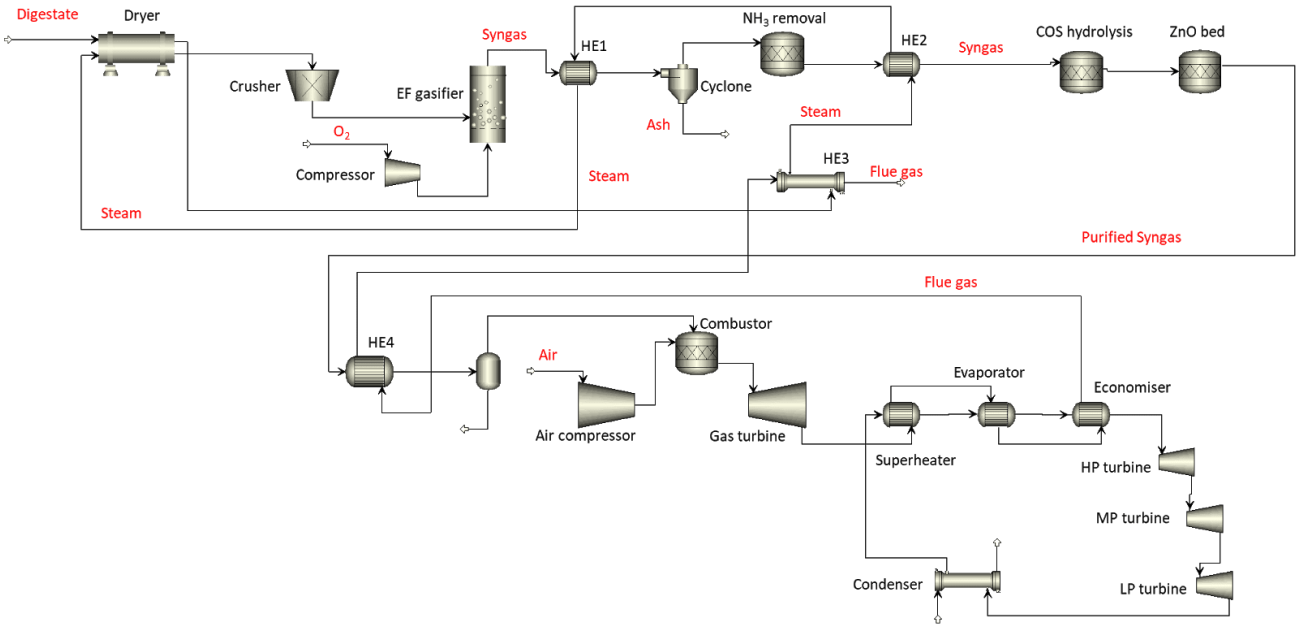


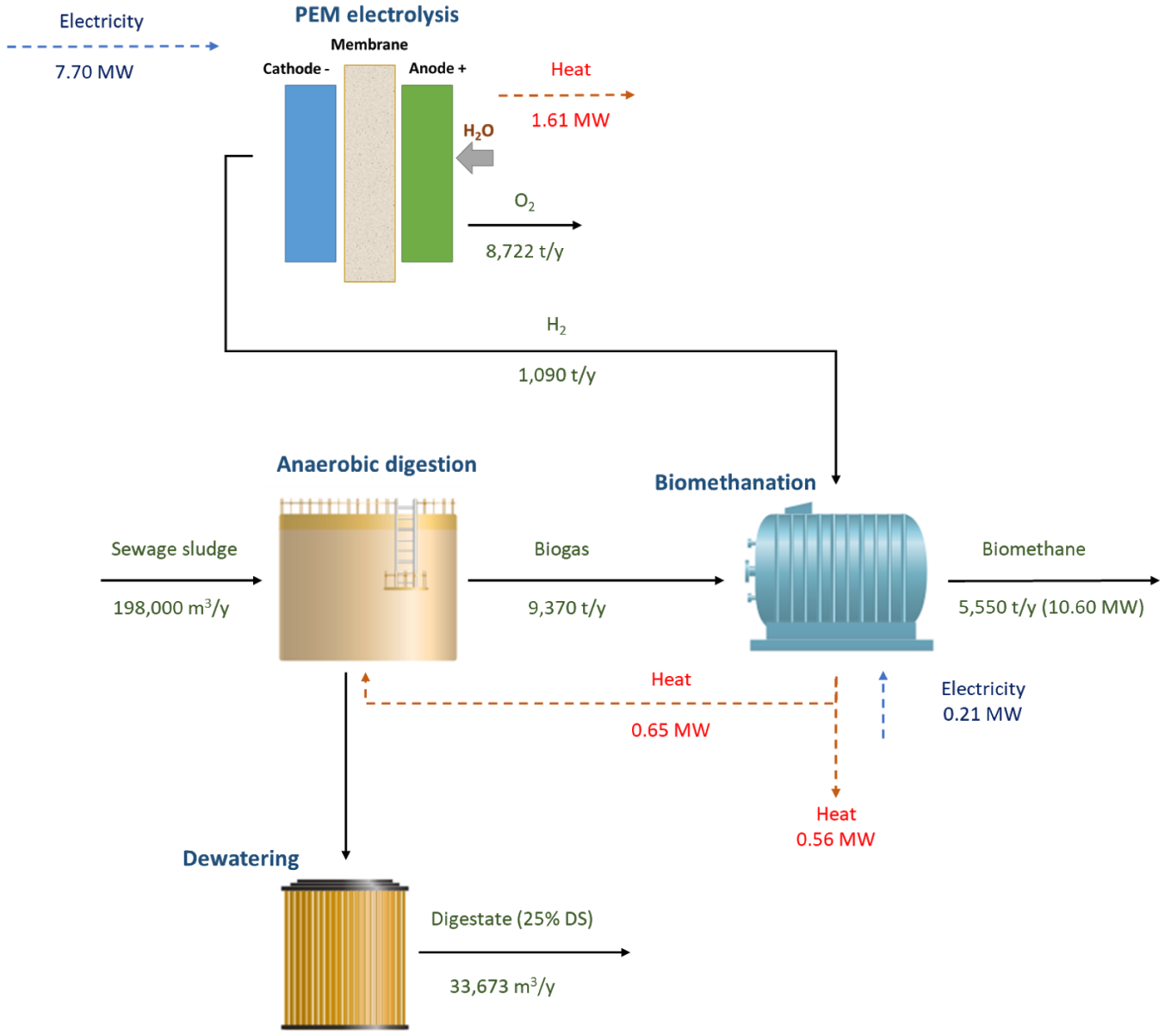
Figure 3. Simplified process flow diagram for the IGCC plant in Scenario 4

4. Results and discussion

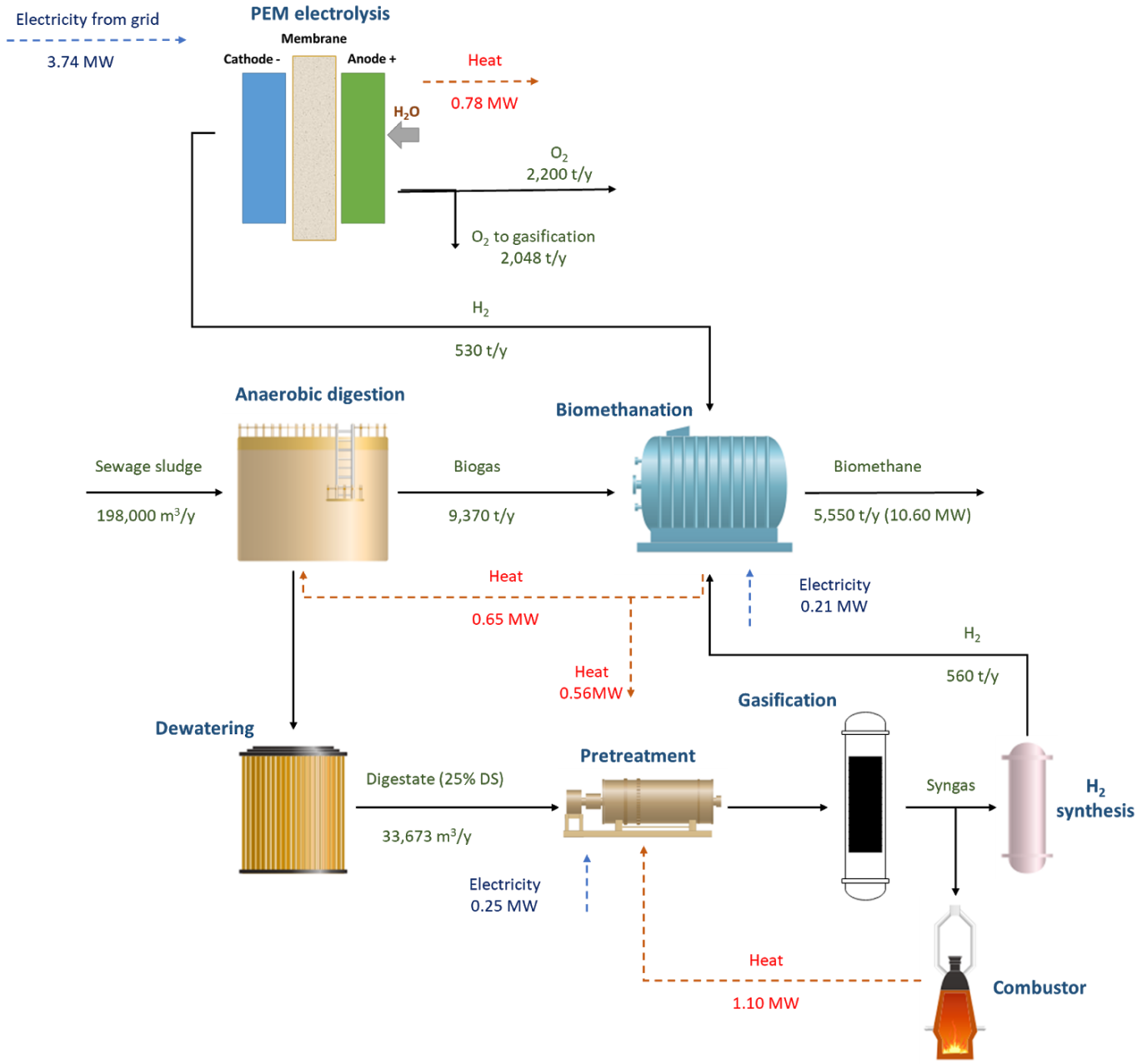
4.1 Mass and energy balances

The mass and energy balances for each scenario are presented in **Table 7** and are also visualised in **Figure 4.** As discussed in Section 2.1, all scenarios treat the same quantity of sewage sludge, while, in contrast to Scenarios 2-4 whereby the digestate is gasified, the digestate in Scenario 1 is safely recycled to farmland (detailed mass balances can be found in the Supplementary material).

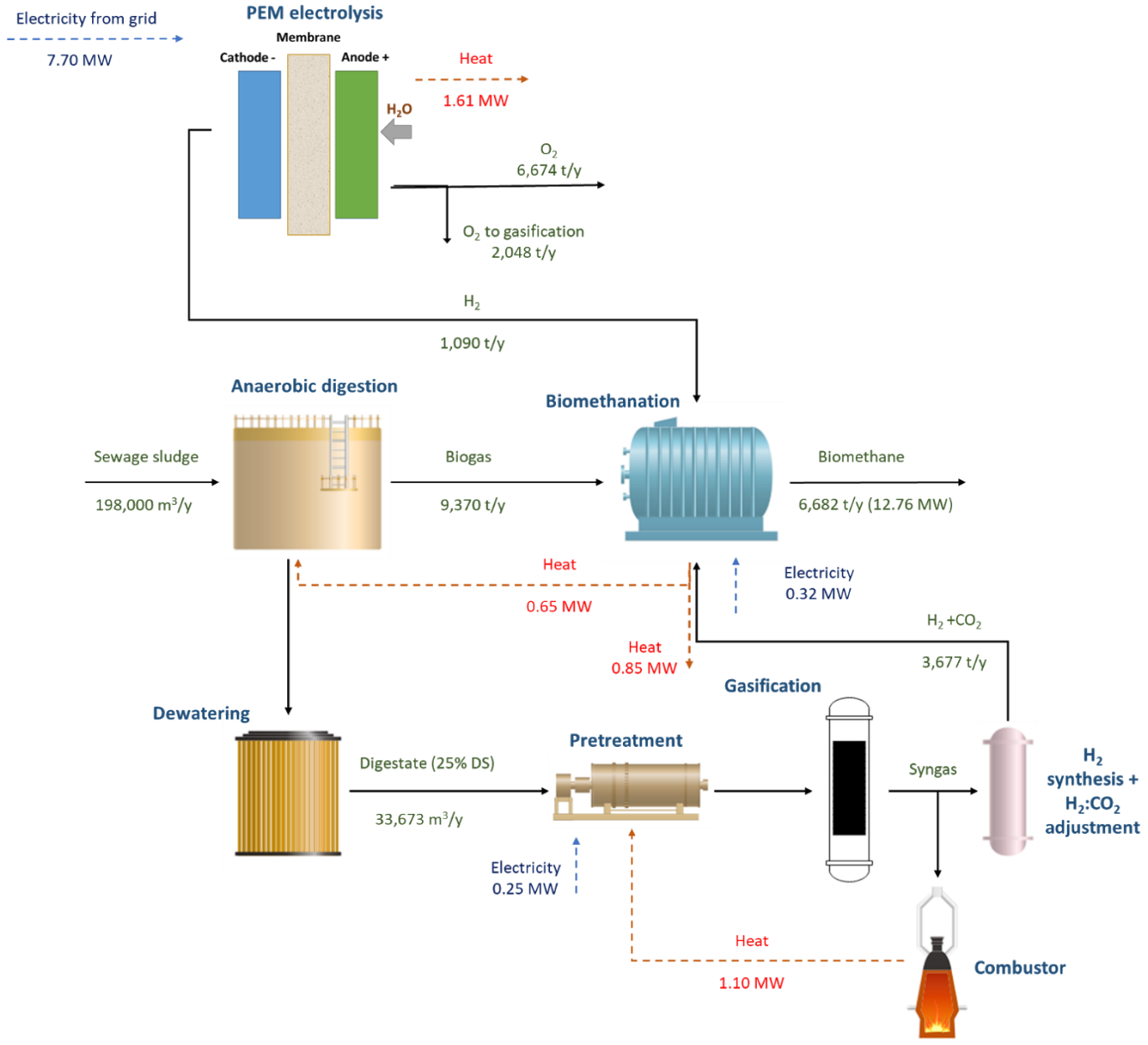
Scenario 1


Electricity
7.70 MW
PEM electrolysis
Membrane
Cathode -
Anode +
Heat
1.61 MW
H_2O
O_2
8,722 t/y
H_2
1,090 t/y
Anaerobic digestion
Biomethanation
Sewage sludge
198,000 m³/y
Biogas
9,370 t/y
Biomethane
5,550 t/y (10.60 MW)
Heat
0.65 MW
Electricity
0.21 MW
Heat
0.56 MW
Dewatering
Digestate (25% DS)
33,673 m³/y

Scenario 2


Electricity from grid
3.74 MW
PEM electrolysis
Membrane
Cathode -
Anode +
Heat
0.78 MW
H_2O
O_2
2,200 t/y
O_2 to gasification
2,048 t/y
H_2
530 t/y
Anaerobic digestion
Biomethanation
Sewage sludge
198,000 m^3/y
Biogas
9,370 t/y
Biomethane
5,550 t/y (10.60 MW)
Heat
0.65 MW
Electricity
0.21 MW
H_2
560 t/y
Heat
0.56MW
Gasification
Dewatering
Pretreatment
Digestate (25% DS)
33,673 m^3/y
Syngas
H_2 synthesis
Electricity
0.25 MW
Heat
1.10 MW
Combustor

Scenario 3


Electricity from grid
7.70 MW
PEM electrolysis
Membrane
Cathode -
Anode +
Heat
1.61 MW
H_2O
O_2
6,674 t/y
O_2 to gasification
2,048 t/y
H_2
1,090 t/y
Anaerobic digestion
Biomethanation
Sewage sludge
198,000 m^3/y
Biogas
9,370 t/y
Biomethane
6,682 t/y (12.76 MW)
Heat
0.65 MW
Electricity
0.32 MW
Heat
0.85 MW
H_2 +CO_2
3,677 t/y
Gasification
Dewatering
Digestate (25% DS)
33,673 m^3/y
Pretreatment
Syngas
H_2 synthesis + H_2:CO_2 adjustment
Electricity
0.25 MW
Heat
1.10 MW
Combustor

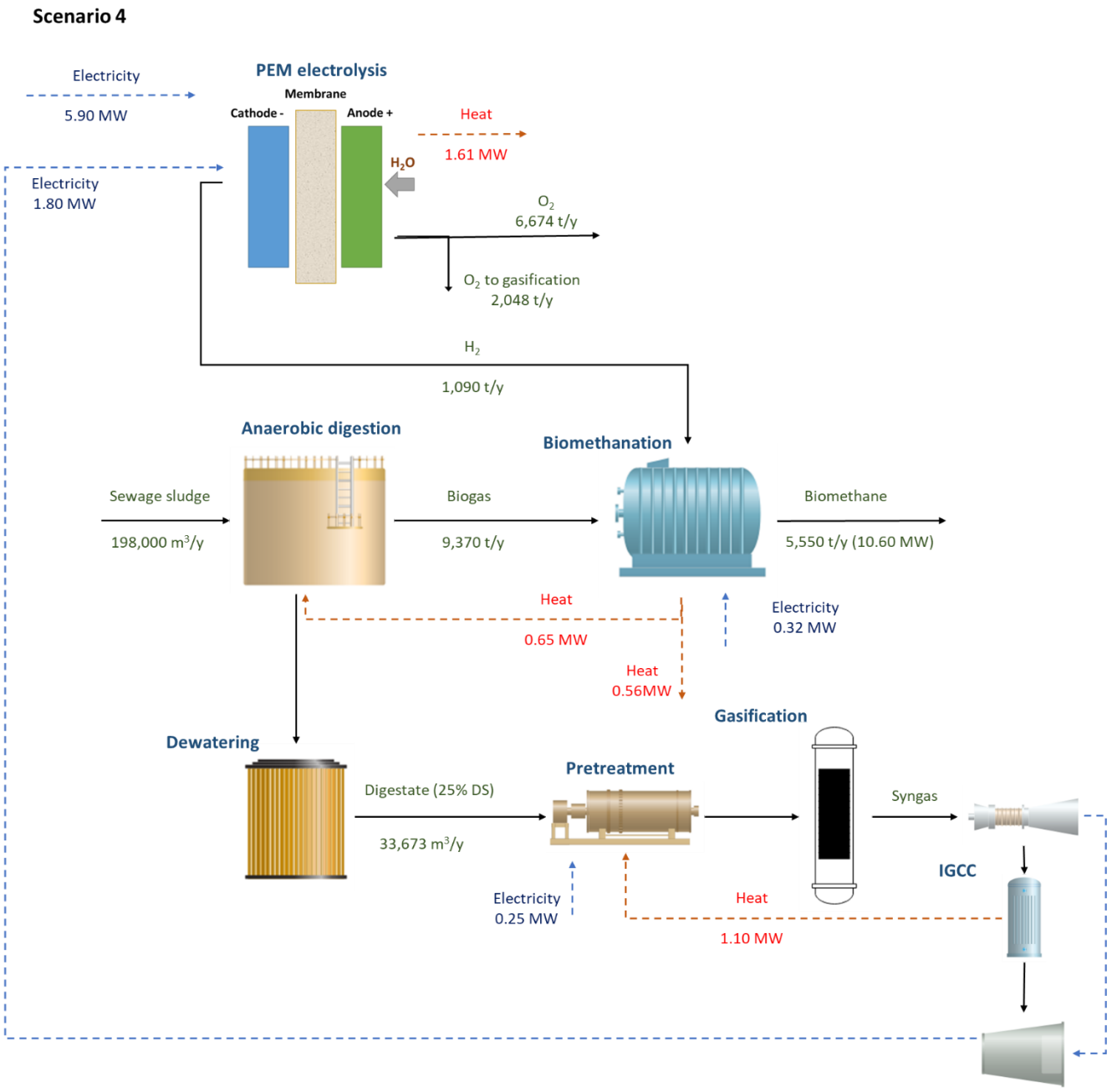


Figure 4. Schematic of the basic mass and energy flows for each scenario investigated.

The electrolyser operates at a constant current density of 2 A/cm^2 that results in a cell voltage of 1.87 V and a stack efficiency of approximately 65% (LHV basis). Furthermore, the consumption of deionised water is taken as 10 L per kg of H_2 [76][77]. As depicted in Eq. (26), the electrolysers were sized based on the H_2 flowrate required for the biomethanation and the stack efficiency.

$$Electolyser\ size\ (MW) = \frac{H_2\ flowrate\ (\frac{kg}{s}) \times LHV_{H2}(\frac{MJ}{kg})}{\eta_{\Delta G}} \quad (26)$$

The electrolyser size for the Scenario 2, *i.e.* 3.4 MW, is perceptibly smaller compared to the other cases, *i.e.* 7 MW, as the H_2 from the gasification plant can meet roughly 51% of the overall H_2 demand of the biomethanation reactor. The Scenario 3 results in increased biomethane yields of approximately 20% compared to the other configurations since the H_2 from the gasification plant (along with CO_2) is now utilised to synthesize more CH_4 in the biomethanation reactor. On the other hand, this design leads to augmented electricity consumption in the biomethanation unit compared to the other scenarios. As expected, the oxygen production follows an identical trend to the electrolyser derived H_2; the Scenario 2 yields the lowest O_2 productivity and the utilised portion, as gasifying agent, is as high as 48%.

As depicted in **Figure 5**, the main electricity consumer for all cases is the electrolyser, being responsible for 72-75% of the overall demand for the Scenarios 1, 3 and 4; assuming that for the Scenario 4 the whole electricity generated in the gasification area supplies the electrolyser. The contribution of the electrolyser for the Scenario 2 is lower compared to the other cases and equal to 57%. It should be noted that in **Figure 5**, the electrolyser includes the system consumption, *i.e.* stack and balance of plant (BoP); the latter accounts for 10% of the stack consumption.

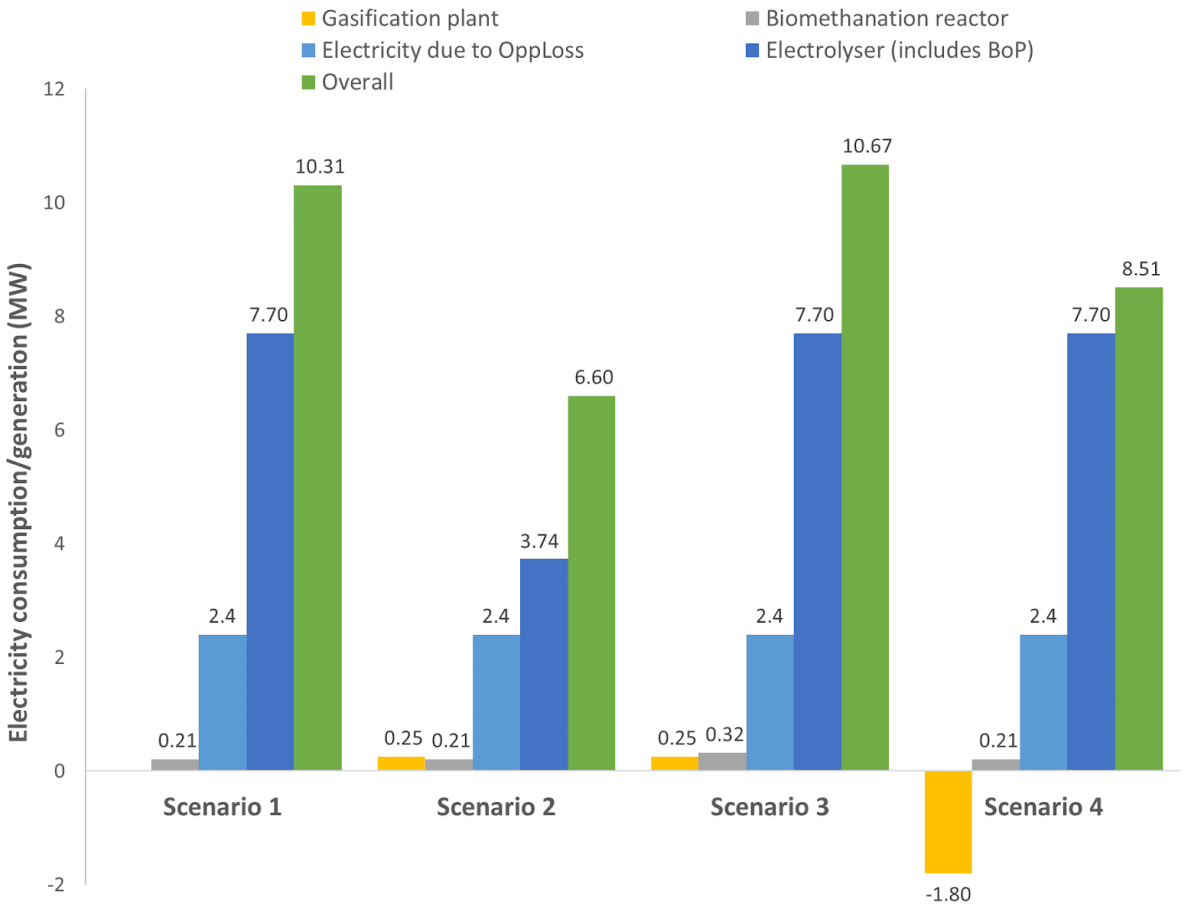


Figure 5. Electricity consumption/generation across the different process segments for each scenario. The minus symbol refers to on-site generation**.**

Heating is needed for all concepts for the operation of the digesters and this duty, *i.e.* 0.65 MW, is covered by heat generated in the biomethanation reactor by means of recirculated water; the generated heat during biomethanation is roughly 1.21 MW and available at 55°C; for the Scenario 3 the available heat is 1.5 MW due to the increased size of the biomethanation reactor. For all cases, the current design does not consider any utilisation for the surplus of heat generated in the biomethanation reactor. In addition, low grade heat (available at roughly 75°C) can be also recovered in the electrolysers but no utilisation for this energy flow was considered in the present study. In theory the quality and quantity of the unutilised heat streams (from biomethanation and electrolysis) can satisfy the demand of at least three more digesters that exist in the WWTP; in this way it might be possible to send more biogas for upgrading (instead to the CHP unit) and the projects to be favoured by the economies of scale.

In addition, for the Scenarios 2, 3 and 4 heating is also required for drying the digestate. On considering the design illustrated in **Figure 2**, the steam that is recirculated between the dryer and the syngas cooling section, absorbs heat internally from the heat exchangers HE3 and HE4 and as no external heating source is required for the entire infrastructure of the Scenarios 2 and 3. Similarly for the Scenario 4, as shown in **Figure 3**, the flue gas is utilised to raise steam in the HE3 and subsequently the steam temperature is increased to 200°C by absorbing heat from the HE1 and the HE2.

Table 7 presents the energy efficiency of each process calculated as the ratio of biomethane energy output (LHV basis) over the sum of the energy content of the sewage sludge and the electricity consumption, Eq. (27). The latter was divided by 0.4 (efficiency of a typical thermal plant) as the other terms in Eq. (27) are expressed in terms of thermal energy. The LHV of the sludge is taken as 14.75 MJ/kg on a dry basis (plant data).

The inclusion of the digestate gasification unit into the design favours the performance of the relevant scenarios, *i.e.* 2-4, from an energetic point view. Based on the energy efficiencies, it can be observed that the most efficient pathway is to reduce the electrolyser's size by introducing H_2 from the gasification area (Scenario 2) rather than keeping the size of the electrolyser identical to Scenario 1 and focus on either increasing the biomethane production (Scenario 3) or generating on-site electricity (Scenario 4).

$$\eta_{plant} = \frac{\dot{m}_{bio-CH_4} LHV_{bio-CH_4}}{\dot{m}_{sludge} LHV_{sludge} + Electricity\ consumed/0.4} \quad (27)$$

Table 7. Mass and energy flows for the investigated scenarios

	Scenario 1	Scenario 2	Scenario 3	Scenario 4
Sewage sludge (m^3/y)	198,000	198,000	198,000	198,000
Digestate after dewatering (m^3/y)	33,673	33,673	33,673	33,673
H_2 from the electrolyser (t/y)	1,091	531	1,091	1,091
Electrolyser stack size (MWe)	7.00	3.40	7.00	7.00
H_2 from the gasification plant (t/y)	-	560	-	-
H_2+CO_2 from the gasification plant (t/y)	-	-	3,677	-
Electricity generation in the gasification plant (MW)	-	-	-	1.8
Oxygen production from the electrolyser (t/y)	8,722	4,248	8,722	8,722
Oxygen utilisation in the gasifier (t/y)	-	2,048	2,048	2,048
Methane production (MW_{HHV} of CH_4)	10.60	10.60	12.76	10.60
Methane production (MW_{LHV} of CH_4)	9.63	9.63	11.60	9.63
Electricity imports (MW)	10.31	6.60	10.67	8.51
*Heating duties for digesters (MW)	0.65	0.65	0.65	0.65
*Heating duties for digestate drying (MW)	-	1.1	1.1	1.1
Energy efficiency (%)	26.5	35.5	31.1	30.2

**Covered internally*

4.2 Economic evaluation

Figure 6 is a thorough breakdown of the purchased equipment cost. The gasification plant notably increases the capital investment for the Scenarios 3 and 4 compared to the Scenario 1 by approximately 53% and 81% respectively. The increase is higher for the Scenario 4 due to the high cost of the gas and steam turbine equipment. For the Scenario 2, the increase in the initial investment compared to the base case is smaller, *i.e.* 12%, as the cost of installing the gasification plant is to a great extent offset by the reduction in the electrolyser cost. The major cost contributor for the Scenarios 1 and 3 is the electrolyser accounting for 74% and 48% of the total costs respectively. For the Scenarios 2 and 4, the gasification plant takes over as the prime cost source accounting for 41% and 45% respectively; nevertheless the electrolyser still poses as a significant expenditure accounting for 36% and 41% for the Scenarios 2 and 4 respectively. Finally, for all cases the share of the H_2S removal unit is negligible.

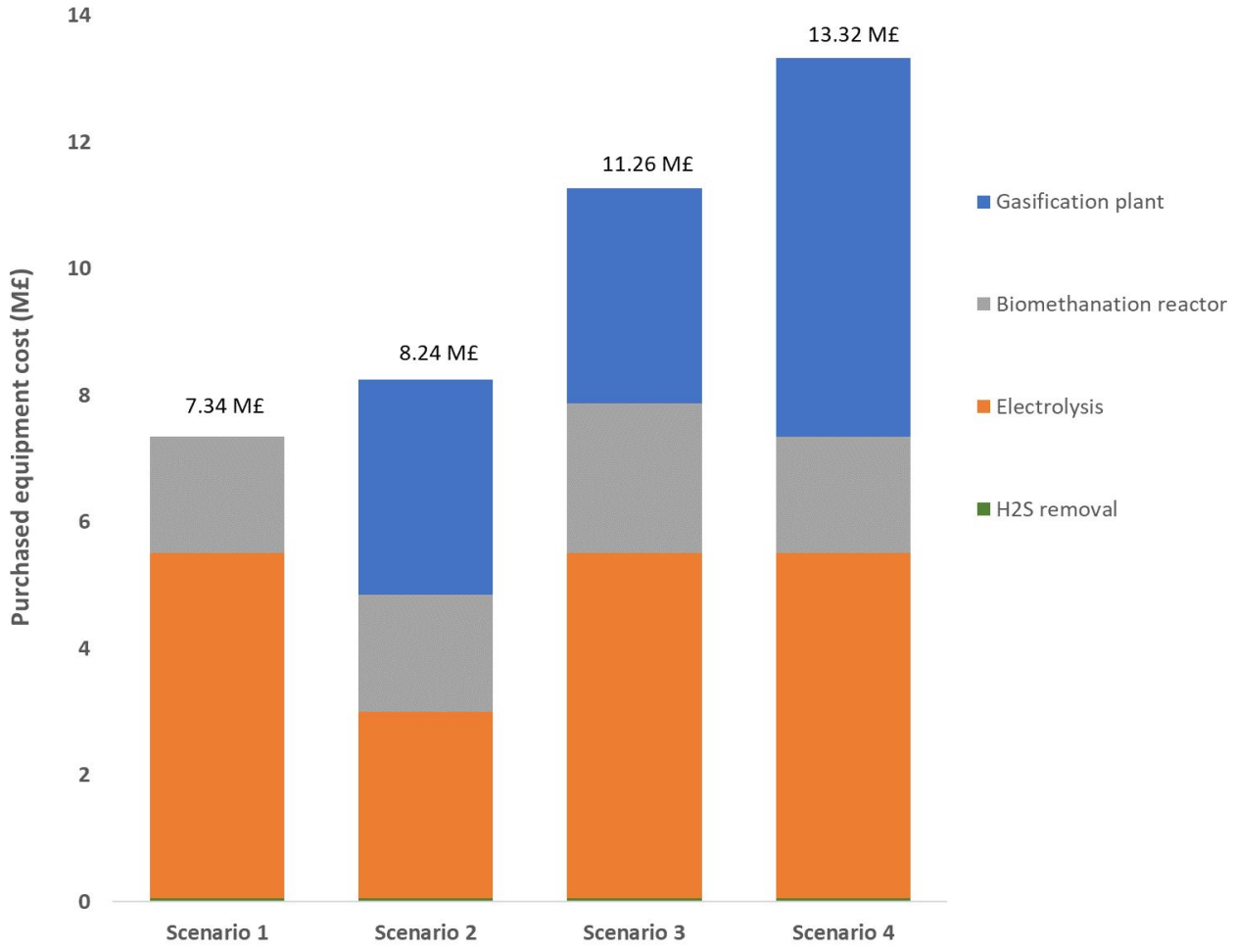


Figure 6. Breakdown of the purchased equipment cost for each scenario. Auxiliary equipment include pumps and the H_2S removal unit

The itemisation of the OPEX for each concept is provided in **Figure 7**. Other variable costs (VC) refer to utilities, catalysts, packing material and ash disposal costs while FOM refers to fixed operating and maintenance costs. The Scenario 3 has the highest OPEX, *i.e.* 12.5 M£, followed by the Scenario 1, 11.01 M£. Electricity clearly dominates the OPEX for all scenarios; approximately 86% for the Scenario 1, 72% for the Scenario 2, 78% for the Scenario 3 and 74% for the Scenario 4. Consequently, the cases with the lower OPEX are those with the less electricity consumption, *i.e.* Scenarios 2 and 4. Therefore, the procurement of cheap electricity is a key factor for the viability of the investigated projects. The FOM contribution is also noteworthy and represents 13%-24% of the overall OPEX.

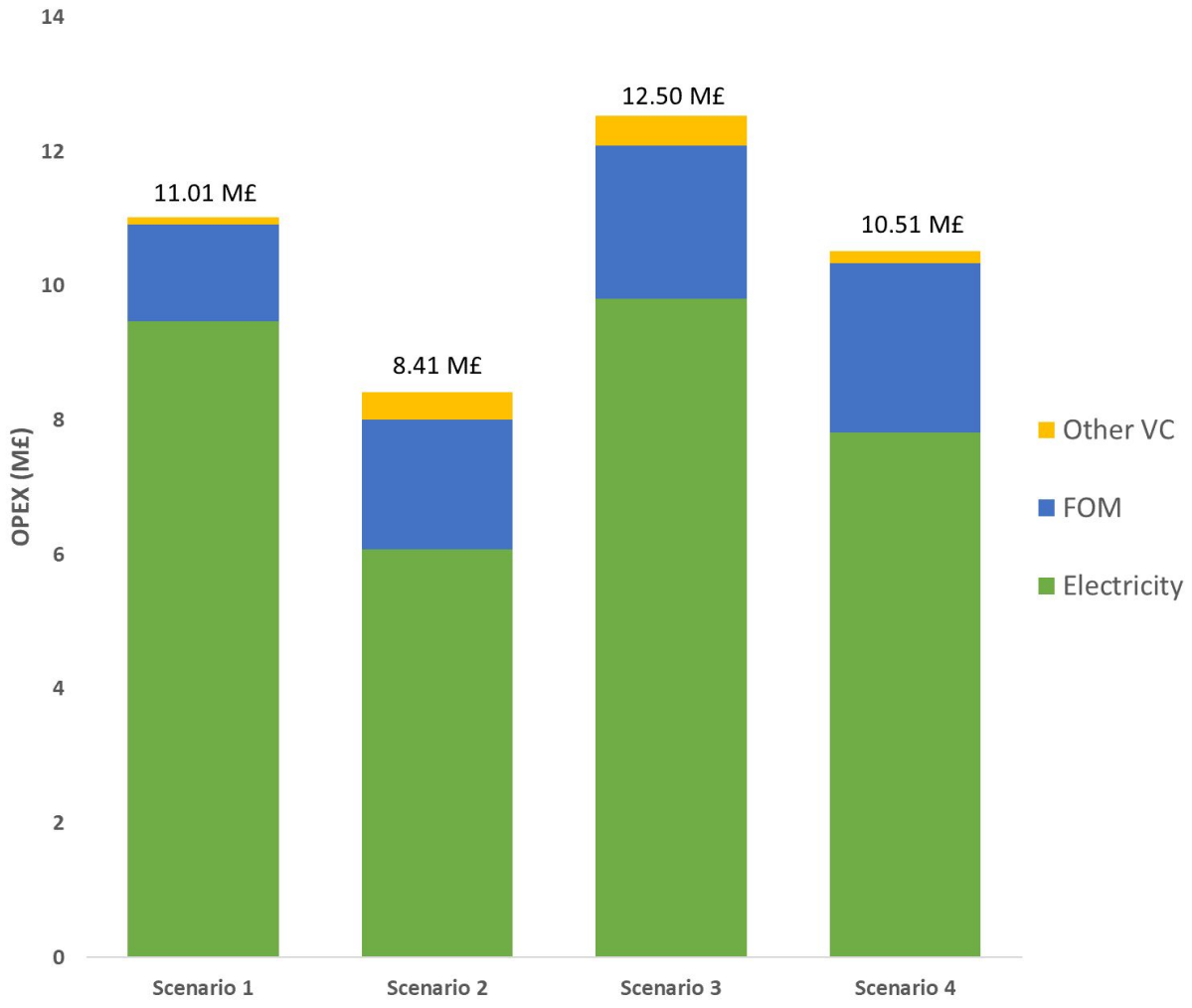


Figure 7. Breakdown of the operating expenses for each scenario

The estimation of the CAPEX and the OPEX set the foundation for the execution of the DCFA. The DCFA resulted in MSPs of 163, 135, 164 and 183 £/MWh (HHV basis) for the Scenarios 1, 2, 3 and 4 respectively. It can be observed that the inclusion of the digestate gasification favours the Scenario 2 while the consideration of implementing the Scenario 4 should most probably be neglected. The Scenario 3 achieved marginally higher selling price than the base case (Scenario 1). However, even if the MSPs are close, it should be noted that the higher initial investment (of Scenario 3 compared to the Scenario 1) carries higher risk and as such it can be discouraging for potential investors. As depicted in **Figure 8**, the MSP for all concepts is more than nine times higher than the NG price, *i.e.* 15 £/MWh (HHV basis) (average 2017)[78], thus suggesting that strong subsidies are required for the commercialisation of the proposed energy systems. Finally, **Table 8** shows the key economic data for each case. It should be also noted that the values of the normalised CAPEX and OPEX indicate that all processes are OPEX intensive.

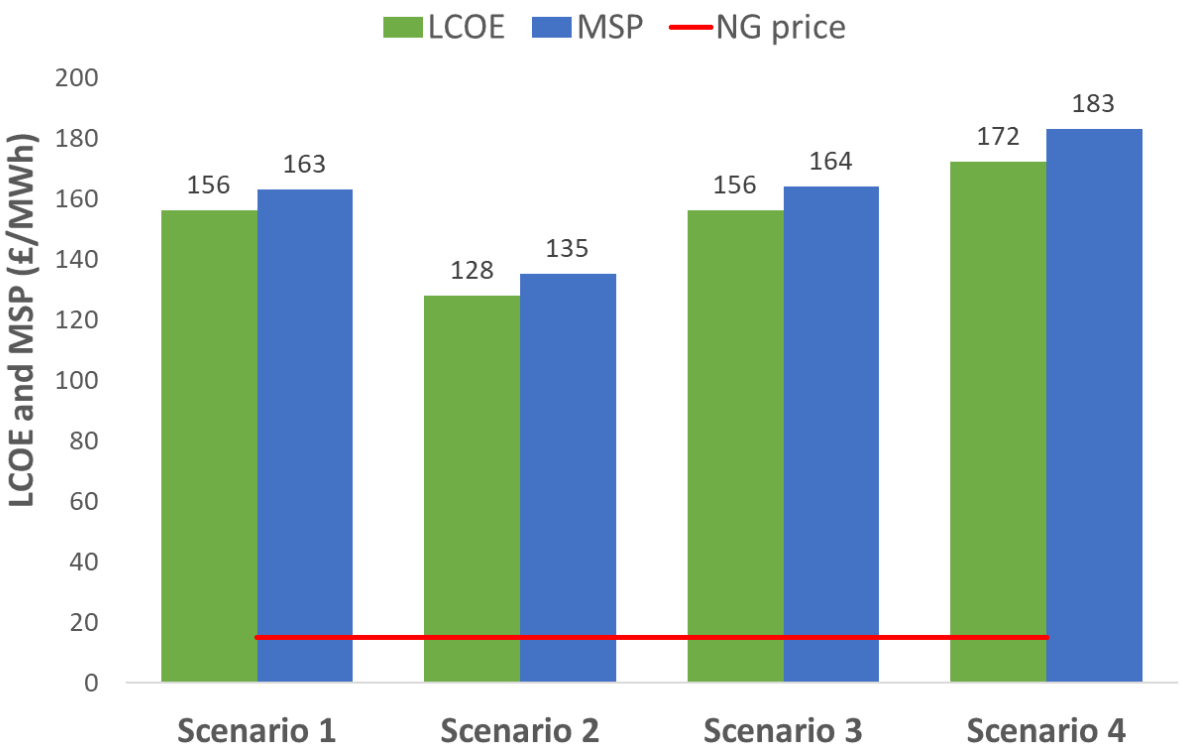


Figure 8. LCOEs and MSPs for each scenario. The NG price (=15 £/MWh) is included for comparisons.

Table 8. Summary of the economic results for each scenario.

	Scenario 1	**Scenario 2**	**Scenario 3**	**Scenario 4**
CAPEX (M£)	19.76	22.16	30.3	35.84
OPEX (M£/y)	11.01	8.41	12.50	10.51
Normalised CAPEX (£/MWh)	27	30	35	49
Normalised OPEX (£/MWh)	129	98	121	123
LCOE (£/MWh)	156	128	156	172
MSP (£/MWh)	163	135	164	183

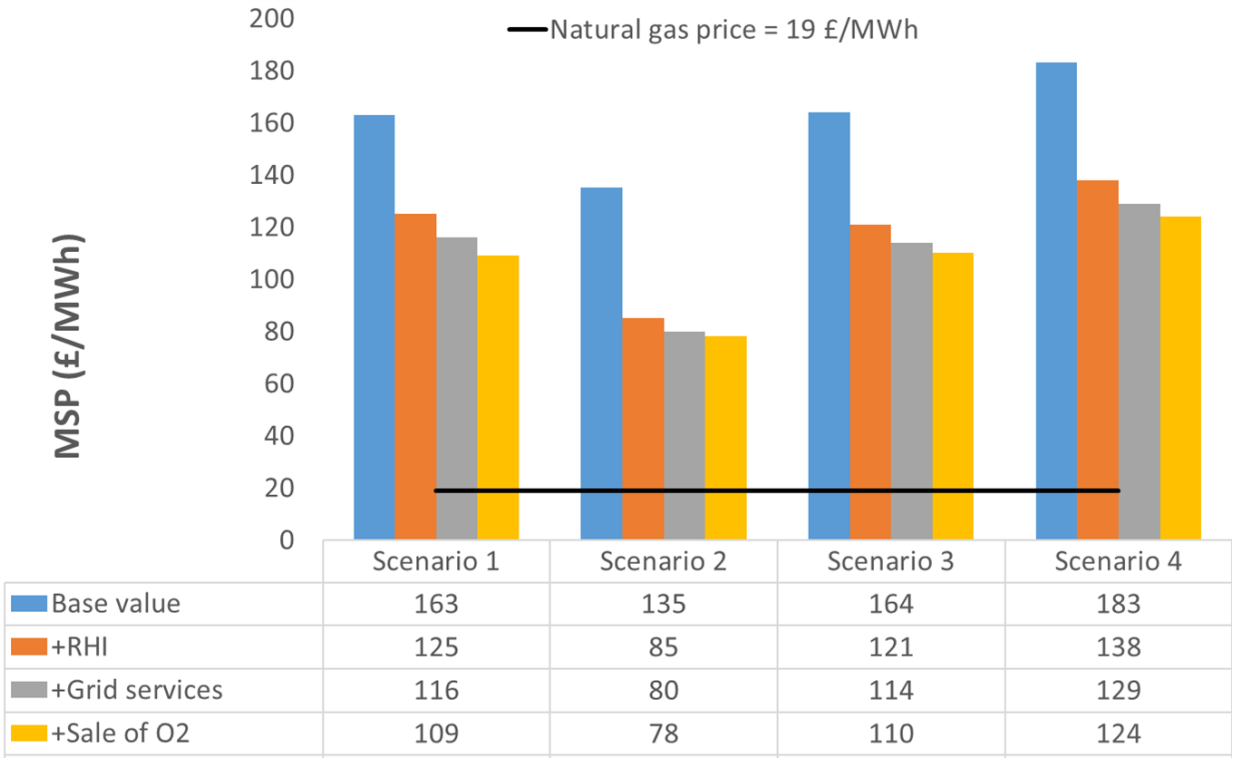
4.3 Effect of policy schemes and by-products credits

The effect of various policy schemes in conjunction with credits received from selling by-products are examined in this section. The renewable heat incentive (RHI) was introduced in the UK in 2009 with the aim of incentivising the generation of renewable heat. The RHI includes the production of biomethane in AD plants and the fee that is paid to the producer for this practise varies based on the biomethane production; the examined WWTP receives a reward of 62 £/MWh (HHV basis). Since for

the base case scenarios electricity is imported from the UK electricity grid, it is reasonable to assume that only the biomethane derived from the AD and the gasification is eligible for any renewable incentive.

Another measure to mitigate the CO_2 emissions is the establishment of a carbon price. Currently, the UK participates in the European Emissions Trading System (EU-ETS) with future considerations to include the establishment of a UK-ETS or the implementation of a carbon tax [79]. The present study assumes a carbon price equal to the present EU-ETS carbon price, *i.e.* ~20£/tonne of CO_2 [80]. The application of the CO_2 price would increase the market price of the fossil NG and as a result the biomethane could be sold at a higher price. On considering an emission factor for the NG of 0.2 tCO_2/MWh (HHV basis) [81], the market price of the NG increases to 19 £/MWh. In addition, the state of the art electrolysers are eligible to receive payments for providing grid balancing services and have the potential to qualify for the Enhanced Frequency Response (EFR) and the Frequency Control by Demand Management (FCDM) markets. A reward of 13.2 £/MWe (installed electrolyser capacity) [82] is assumed in this study to account for the grid balancing services. An additional advantage of the electrolysers is the production of high purity oxygen. Oxygen has a relatively high market value due to its demand in steel and chemical industries [83]; a market price of 70 £/tonne of O_2 [84] is considered herein.

As depicted in **Figure 9**, the reduction of the MSP for each case is in the range of 32%-42%. The highest percentage drop is witnessed in the Scenario 2 (MSP = 78£/MWh). The Scenarios 1, 3 and 4 experience similar reductions of 32%-33%. Also, it can be observed that the effect of the RHI is much more significant than the other benefits. The effect of RHI is more significant on the Scenarios 2, 3 and 4 compared to the Scenario 1 as in the latter the whole amount of the additional biomethane is produced via electrolysis while for the former cases digestate gasification also contributes to the production of the hydrogenated biomethane. At a carbon price of 20 £/t CO_2, the MSP of the Scenarios 1, 2, 3 and 4 are 5.7, 4.1, 5.8 and 6.5 respectively times higher than the NG price. It is clear that further incentives are required or techno-economic improvements have to be achieved to enhance the competitiveness of the proposed infrastructures. An additional payment that ranges between 59 and 105 £/MWh is required.



	Scenario 1	Scenario 2	Scenario 3	Scenario 4
Base value	163	135	164	183
+RHI	125	85	121	138
+Grid services	116	80	114	129
+Sale of O2	109	78	110	124

Figure 9. Cumulative effect of various policy schemes and O_2 sale credits on the MSP for each scenario. The NG price, for a carbon price of 20 £/t CO_2, is illustrated for comparisons.

4.4 Sensitivity analysis

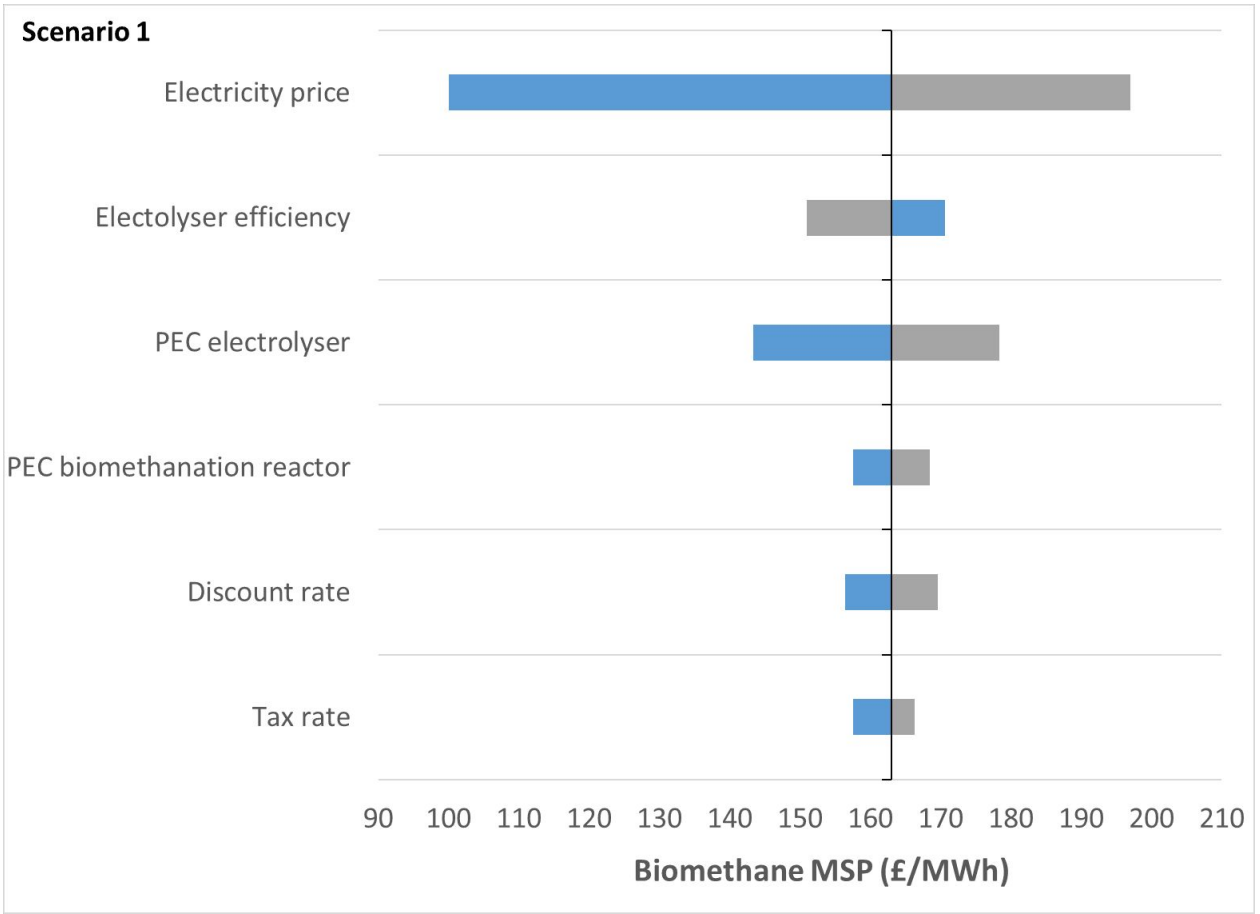
Sensitivity analysis determines how a measure of worth is altered when one or more parameters vary over a selected range of values. Typically one parameter at a time is varied, and independence with other parameters is assumed. The investigated parameters along with their upper and lower limits are presented in **Table 9**. Based on the techno-economic analysis presented above, the electricity price is the dominant OPEX while the purchased equipment cost of the electrolyser and the gasification plant monopolise the CAPEX. In addition, the effect of the cost of the core unit, *i.e.* the biomethanation reactor, was investigated in conjunction with the efficiency of the electrolyser since it directly affects the electricity consumption. Finally, typical market relevant factors, such as the tax rate and the discount rate, were also selected. The effect of the chosen variables on the MSP for each scenario is illustrated in **Figure 10**.

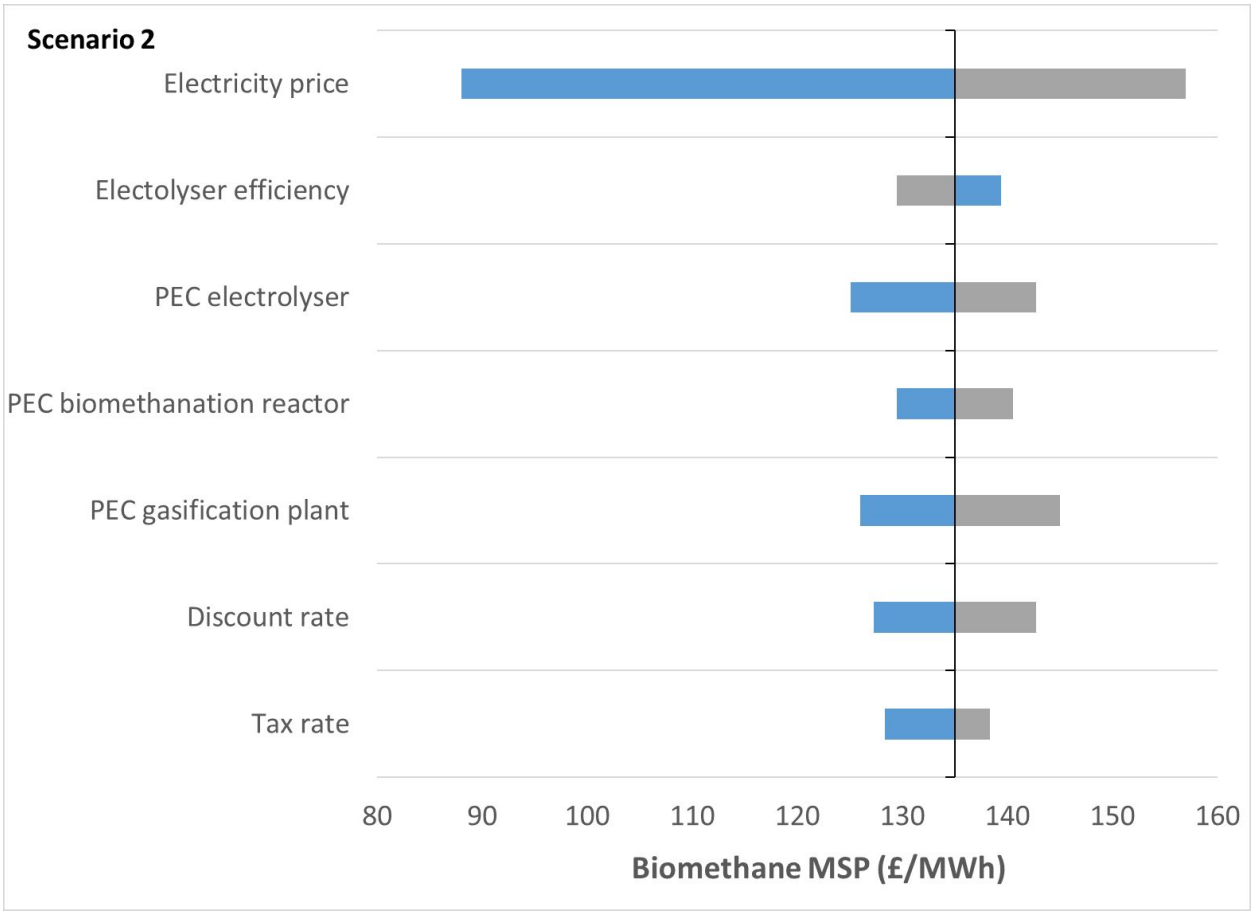
For all the cases investigated, the electricity price raises the greatest uncertainty. If the electricity price drops to 50 £/MWh, then the MSP decreases by 39%, 35%, 38% and 34% for the Scenarios 1, 2, 3 and 4 respectively; in an optimistic scenario such a low electricity price could be in theory achieved by on site renewable electricity generation [85]. Another significant source of uncertainty for the Scenarios 1 and 3 is the electrolyser equipment cost. For the Scenarios 2 and 4 the impact of the electrolyser cost is roughly equal to the equipment cost of the gasification plant. In addition, for all

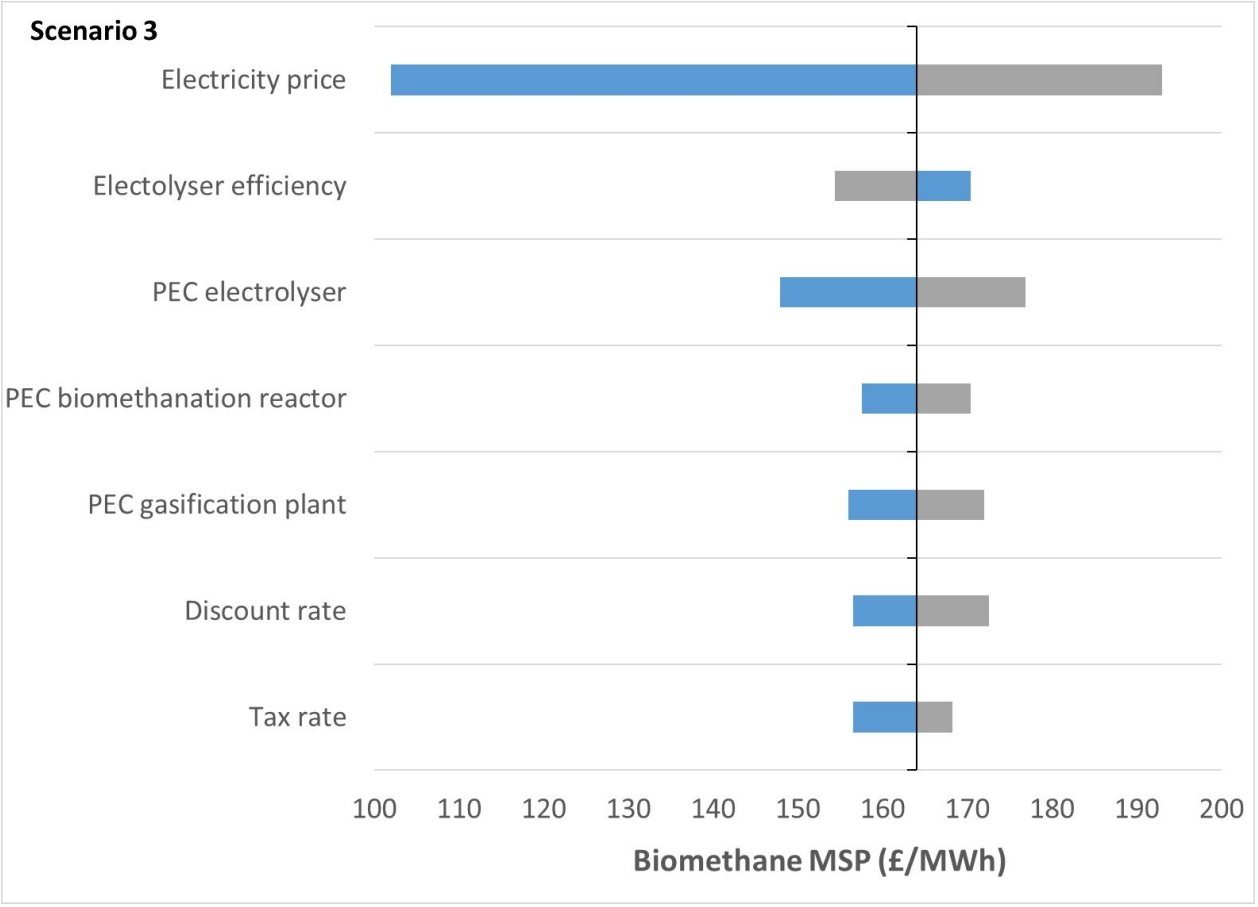
scenarios, the biomethanation reactor has a relatively low influence on the respective MSPs; a similar effect is observed for the discount rate. The equipment cost of novel technologies can fall in the future due to learning and economies of scale effects; this is the case for PEM electrolysis units as argued in [86] and for the biomethanation unit the successful operation of two demo plants [16] can pave the way for further development of this technology. Further, all concepts are less sensitive to the tax rate, indicating that efforts on improving the endogenous techno-economic variables should be prioritised.

Table 9. Chosen parameters for sensitivity analysis and their bounds.

Parameter	Low value	Nominal	High value	Unit
Electricity price	50	115	150	£/MWh
Electrolyser efficiency	60	67.2	75	%
PEC electrolyser	0.45	1.044	1.5	M£/MW
PEC biomethanation reactor	1	1.97	3	M£/5 $MW_{th,CH4}$
PEC gasification plant	50	100	150	%
Discount rate	8	10	12	%
Tax rate	0	30	40	%




Scenario 2
Electricity price
Electolyser efficiency
PEC electrolyser
PEC biomethanation reactor
PEC gasification plant
Discount rate
Tax rate
80
90
100
110
120
130
140
150
160
Biomethane MSP (£/MWh)


Scenario 3
Electricity price
Electolyser efficiency
PEC electrolyser
PEC biomethanation reactor
PEC gasification plant
Discount rate
Tax rate
100 110 120 130 140 150 160 170 180 190 200
Biomethane MSP (£/MWh)

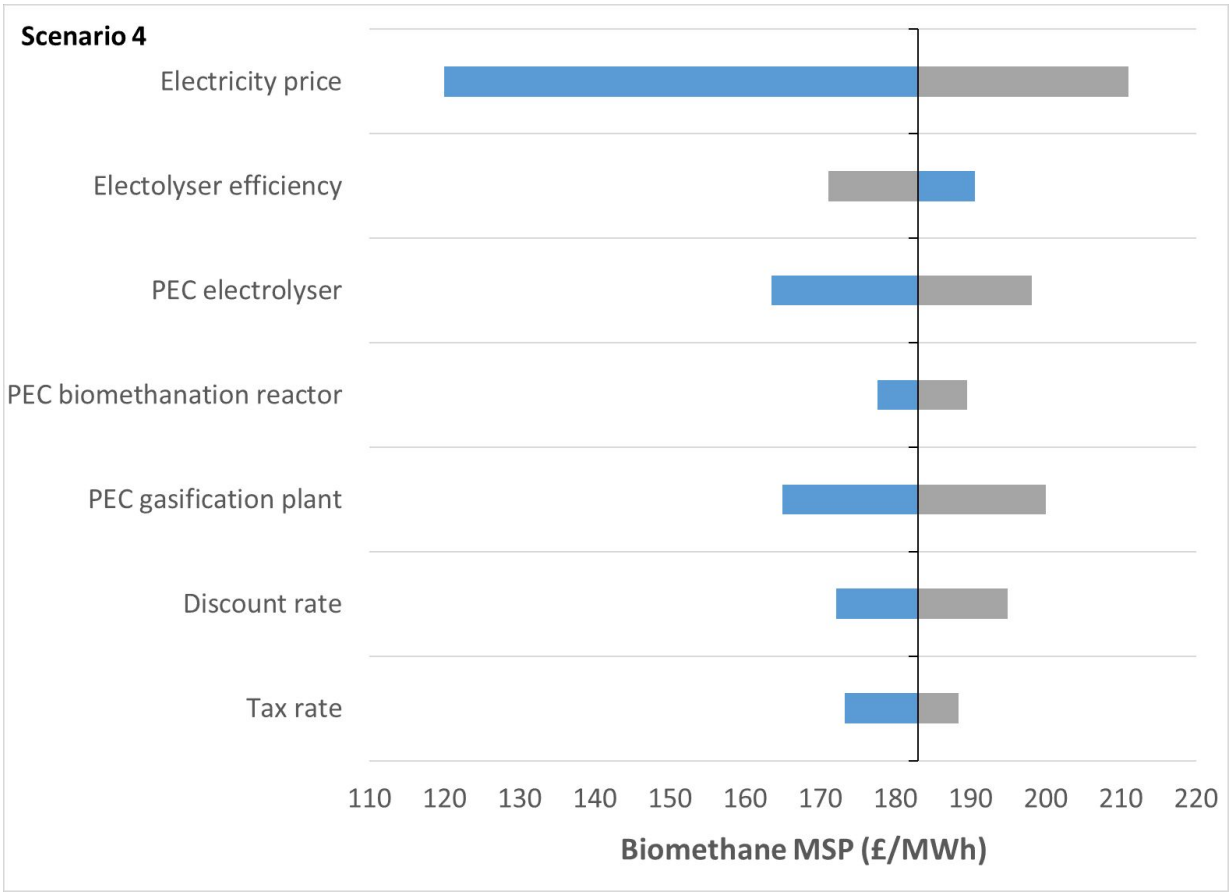

Scenario 4
Electricity price
Electolyser efficiency
PEC electrolyser
PEC biomethanation reactor
PEC gasification plant
Discount rate
Tax rate
110 120 130 140 150 160 170 180 190 200 210 220
Biomethane MSP (£/MWh)

Figure 10. Sensitivity analysis on the MSP for each scenario

4.5 Business cases for 2030

This section delivers a prospective scenario analysis of the examined concepts. Scenario analysis is the process of considering scenarios for evaluating potential future events. Scenarios are alternative, though not equally likely states of the world that represent plausible conditions under different assumptions [87].

The techno-economic assessment, presented in Sections 4.1 and 4.2, confirmed that the biomethanation technology is not competitive under the current market and technical conditions. Nevertheless, the chief cost drivers are associated with the electrolysis, *i.e.* electricity price and CAPEX, and not the biomethanation. Several advances are projected that have the potential to drastically decrease the cost of the electrolyser [86]. These developments include operating window optimisation (*e.g.* temperature, current density), economies of scale, lower cost materials and manufacturing savings (*e.g.* reduced production time) [86]. In order to attain lower electricity prices, on-site renewable energy generation should be realised; for example a dedicated wind farm that would solely supply electricity to the proposed systems. Similar infrastructures are already under operation such as a hydrogen fuel station in South Yorkshire operated by ITM, whereby a 225kW wind turbine is coupled directly to an electrolyser [88]. Thus, we assume on-site electricity generation from onshore wind for the 2030 scenarios. The levelised cost of electricity was taken as 60 £/MWh [85].

Table 10 summarises the techno-economic features that are projected to be actualised by 2030. Based on these aspects, the biomethane MSP for each scenario was calculated for 2030. It should be highlighted that probable cost reductions in the gasification technology, due to learning effects, were not considered as there are currently only a couple of demo plants that produce H_2 from biomass [89] and there are no plans for the deployment of new plants by 2030 [90][91]. In addition, no cost reduction was assumed for the biomethanation reactor so as to neglect uncertainties related to the definition of the progress ratio and the assumption of the number of any new plants [92] deployed by 2030.

Furthermore, the effects of the RHI and the carbon price were also included. Renewable incentives for biomethane production will most probably continue to be effective in the future with the aim of decarbonising the heat sector as the consumption of natural gas is anticipated to grow from 120 trillion cubic feet (Tcf) in 2012 to 203 Tcf in 2040 [93]; also, it should be noted that the current market driver for biogas upgrading to biomethane, regardless the deployed upgrading technology, is the RHI. In addition, since renewable electricity is utilised in the 2030 scenarios, it was assumed that

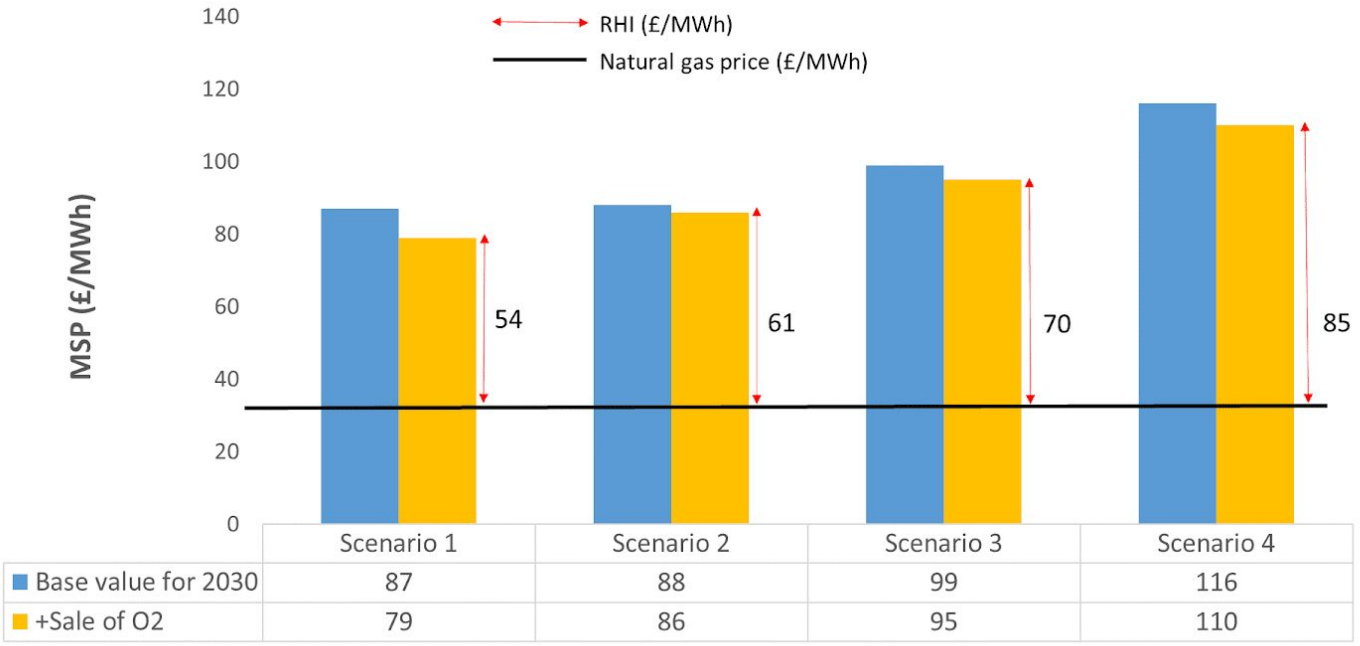
the whole amount of the biomethane produced (from AD and biomethanation) is eligible for renewable incentives. As it is hard to forecast the size of future renewable incentives we calculated herein the RHI values for each scenario that will match the MSPs with the NG price.

Regarding the carbon price, developed models have calculated that the carbon price should be notably increased to 55 €/tCO_2 (~50£/tCO_2) in order to meet the Paris Agreement targets [94]; hence this value was adopted in the present study and the resultant NG price is 25 £/MWh. Furthermore, as on-site electricity generation is assumed, the projects are not eligible any more for receiving grid balancing fees.

Table 10. Projected values of chosen techno-economic parameters in 2030.

Parameter	Low value	Unit	Reference
Electricity price	60	£/MWh	[85]
Electrolyser efficiency	70	%	[95][96]
PEC electrolyser	0.45	M£/MW	[95][97]
Carbon price	50	£/tCO_2	[94]

Figure 11 depicts the projected MSPs for each case investigated in 2030 along with the effect of several policy mechanisms. The cases that are more dependent on the electrolysis unit (greater installed capacity) experience the greater relative reductions in the base case MSP compared to the respective 2017 MSPs. As such, a greater fall is observed in the Scenario 1, *i.e.* 47%, followed by the Scenario 3, *i.e.* 40%, and the Scenario 4, *i.e.* 37%, while the lowest drop, *i.e.* 35%, is detected in the Scenario 2.



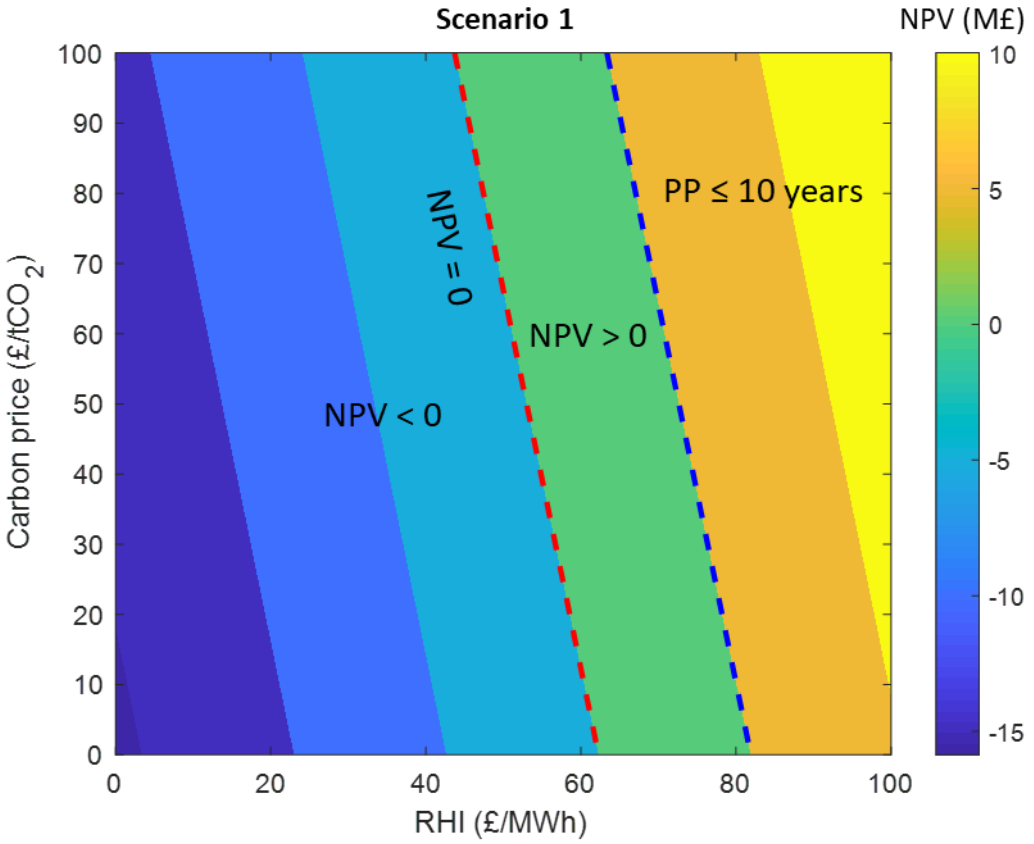
	Scenario 1	Scenario 2	Scenario 3	Scenario 4
Base value for 2030	87	88	99	116
+Sale of O2	79	86	95	110

Figure 11. Projected MSPs for each case investigated in 2030 that includes incomes from selling oxygen and the required value of the RI to match the NG price. The NG price has been calculated for a carbon price of 50 £/t CO_2 and it is 25£/MWh.

After the implementation of the carbon price and the revenues received from O_2 sale, it was revealed that the necessary values of the RHI for the MSPs to break-even with the NG price are 54, 61, 70 and 85 £/MWh. The required RHIs in 2030 for the Scenarios 1 and 2 are lower than the existing RHI, signifying that there are good prospects for profitability in the future for these projects. By comparing the gasification based scenarios, it is apparent that the Scenario 2 outplays the Scenario 3 while the 2030 techno-economic conditions do not help the Scenario 4 to improve against the competition, exhibiting that the installation of an IGCC unit remains uncompetitive in terms of economic and energetic performance. Thus, in the case that the gasification plant is deployed in the foreseeable future, it seems to be preferable to target hydrogen generation (Scenario 2) and reduce the size of the installed electrolyser. It appears, therefore, that the Scenarios 1 and 2 hold greater potential.

In addition, we examined the effect of the RHI and the carbon price on the profitability of the Scenarios 1 and 2. **Figure 12** depicts the economically feasible regions for the Scenarios 1 and 2 with respect to the RHI and the carbon price. The red line refers to the breakeven cases while the blue line to the cases that generate profits at reasonable payback period (~10 years); left to the red line the NPV is negative while right to the blue line profitable scenarios exist with payback periods less than 10 years. It is observed that for the Scenario 1 there are more opportunities for financial success; the area of the positive NPV region is greater than for the Scenario 2. In the case that the

RHI is 62£/MWh (the current price of the RHI) and no carbon price is applied, the Scenario 1 reports profits of 0.4 M£ whereas the Scenario 2 losses of 2.6 M£. A carbon price of 50 £/tCO_2 (and RHI=62£/MWh) would result in positive NPVs for both Scenarios, i.e. 2.3 M£ for the Scenario 1 and 0.1 M£ for the Scenario 2. In order to achieve payback periods less than 10 years, for carbon price of 50 £/tCO_2, the RHI should increase to 72£/MWh and 82 £/MWh for the Scenarios 1 and 2 respectively. Finally, for a zero carbon price scenario, the projects can break-even with the NG price for an RI of 64 £/MWh (Scenario 1) and 71 £/MWh (Scenario 2).



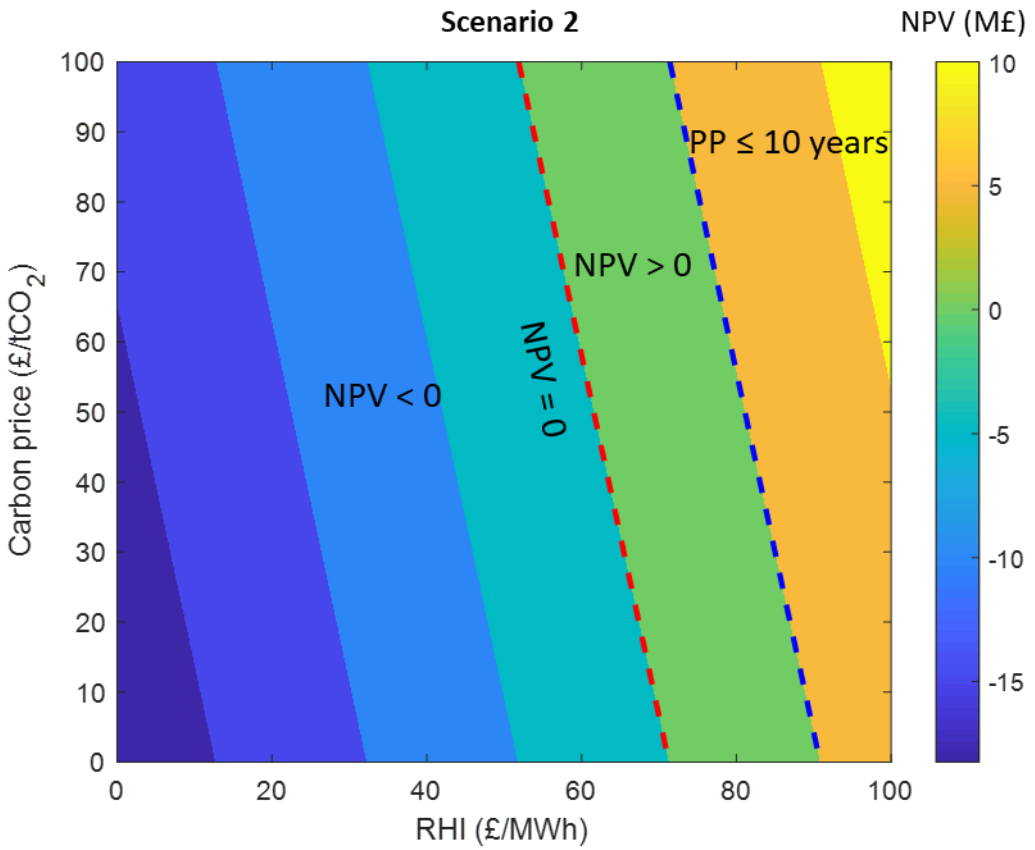


Figure 12. NPV for the 2030 Scenarios 1 and 2 as a function of the RHI and the carbon price

Overall, provided that the estimated advances in the electrolyser will be accomplished and on-site generation of electricity will not be a major issue, the Scenario 1 is the most cost-effective option. The deployment of the gasification unit does not add appear to add significant value to the biomethanation concept and it seems that other methods should be considered for the utilisation of the digestate in the short-term. It appears that the establishment of less complex renewable energy systems should be prioritised. The inclusion of additional technologies increases the capital expenditures and the project risk and as such investment in such ventures is less probable. Nevertheless, the inclusion of digestate gasification in the mid- to long-term is a promising solution to achieve circular economies and therefore prompt R&D support can accelerate its deployment.

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5. Concluding remarks

The study details a comprehensive techno-economic assessment of four power to gas systems based on biomethanation and gasification technologies. Modelling approaches were applied to quantify mass and energy flows and subsequently an economic model was developed to explore the feasibility of the proposed energy infrastructures.

The energy efficiency of the examined scenarios range from 26.5% to 35.5%; the lower limit corresponds to the standalone biomethanation concept (Scenario 1) as the digestate remains unexploited. The cost of hydrogen production was identified as the main cost driver and depending on the scenario, this includes electricity price, the cost of the electrolyser and/or the cost of the gasification plant. The scenario that couples AD with the digestate gasification, targeting at hydrogen production (Scenario 2), achieves the lowest MSP, *i.e.* 135 £/MWh, followed by the Scenario 1 (MSP=163£/MWh); the Scenario 1 suffers from increased electricity consumption when compared to the Scenario 2. The Scenario 3 that targets to enhance the biomethane yield by feeding the biomethanation reactor with a H_2/CO_2 gas stream (derives from the gasification plant) attains an MSP, of 164 £/MWh. The deployment of an IGCC unit (Scenario 4) is clearly the least cost-effective alternative (MSP=183 £/MWh) and this is mainly due to the significantly increased capital investment compared to the Scenario 1 and the lower energy efficiency compared to the other gasification based cases.

The provision of monetary rewards, such as the RHI and a fee for providing grid balancing services along with credits by selling oxygen, causes a reduction in the MSPs of 32%-42%. The sensitivity analysis revealed that all concepts are primarily OPEX intensive since they are more sensitive to the electricity price and secondarily CAPEX intensive with the equipment cost of the electrolyser and the gasification plant raise notable uncertainty. Other influential factors include the electrolyser efficiency and the discount rate.

Based on the current market status none of the scenarios are competitive with the NG even after the incorporation of incentives and the establishment of a carbon price. Hence, a 2030 analysis was also carried out that considers projected equipment cost reductions and on-site electricity generation. This scenario analysis exposed that the establishment of long-term renewable incentives is a key factor for profitability. On considering that the 2030 RHI will be in parity with the 2017 RHI the Scenario 1 can generate profits even without the implementation of a carbon price. The Scenario 2 can be also competitive but the establishment of an appropriate carbon price is a requirement for economic viability. The Scenarios 3 and 4 appear not to be competitive even under the techno-economic conditions in 2030. The prospects for the Scenario 1 seem to be greater as its dependency on the cost of producing hydrogen is much higher than the other cases (where gasification poses also as a significant cost driver) and as such it benefits more from the reduced electricity price and the anticipated electrolysis advances.

Collectively, the future of biomethanation is directly linked with the future of water electrolysis. The progress of these technologies should go hand in hand as they can benefit from each other; the

conversion of hydrogen to a more flexible substance can foster the expansion of power to gas systems and justify investment in R&D. The adoption of a circular economy approach showed promising results but more complicated systems increase the investment risk. It appears to be more sensible for the upcoming decade to concentrate efforts on establishing an efficient stand-alone P2G sector, with infrastructures similar to the Scenario 1, and subsequently integrated energy systems can be considered.

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Appendix A – Economic data and methodologies

Table A.1. Methodology for fixed costs [43][44][45]

Parameter	Price
Supervision	0.25×Labour
Direct overhead	0.5×(Labour + supervision)
General overhead	0.5×(Labour + supervision + direct overhead)
Maintenance Labour	0.015×FCI
Maintenance materials	0.015×FCI
Insurance and tax	0.01×FCI
Replacement cost of the electrolyser (% of installed capital cost; applied after 10 years)	12
Financing working capital	Discount rate × WC

Table A.2. Methodology for variable costs

Parameter	Price	Unit	Reference
Electricity	115	£/MWh	Plant data
Digestate transport to farm land	10	£/t	Plant data
Cooling water	0.025	£/t	[43]
Deionised water	7.64	£/t	[46]
Ash disposal	21.1	£/t	[47]
Nickel catalyst	24	£/t	[48]
COS catalyst	1,797	£/m^3	[49]
WGS catalyst	13,836	£/m^3	[49]

PSA packing	0.92	£/kg	[50]

Where necessary cost were converted to GBP and updated considering an average yearly inflation rate of 2%. Catalyst are replaced every 3 years.

Table A.3. Equipment cost data

Equipment	Base cost (M£)	Design Variable	Base capacity	Unit	Scaling factor
Biomethanation reactor	1.97	$MW_{th,CH4}$	5	MW	0.7
Pretreatment*	10.65	Biomass input	2000	dt/d	0.72
Gasification**	31.7	Biomass input	2000	dt/d	0.72
WGS reactor	1.39	Total gas feed	150	kg/s	0.67
Cyclone	0.04	Total gas flow	1	m^3/s	0.7
ZnO guard bed	0.016	Gas flow	8	m^3/s	1
Burner	1.576	Heat duty	20	MW	0.83
PSA	4.71	Purge gas flow	0.294	kmol/s	0.74
Compressors	0.39494	Power	413	kW	0.68
Heat exchangers	0.20956	Area	1000	m^2	1
Pumps	0.0806	Power	10	m^3/s	0.36
Steam turbine	0.27404	Power	10.5	MW	0.44
NH_3 removal unit	46.85	Syngas flow	31,000	kmol/h	0.9
Electrolysis	1.044	Installed capacity	1	MWe	0.85

*Pretreatment includes dryer and grinder, **Gasification includes gasifier, lockhopper and syngas coolers