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TECHNO-ECONOMIC OPTIMISATION FOR OPTIMAL CONCEPTUAL CHP PLANT DESIGN WITH HEAT EXCHANGERS AND BOP

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GLOSSARY

A_{ex}	Heat exchanger area (m^2)
AMPL	A Mathematical Programming Language
AOG	Anode off-gas
BLAZE	Biomass Low cost Advanced Zero Emission small-to-medium scale integrated gasifier-fuel cell combined heat and power plant
CAPEX	Capital expenditure (MEUR)
CGE	Cold gas efficiency
CHP	Combined heat and power
COG	Cathode off-gas
DBFBG	Double bubbling fluidised bed gasifier
Eff_{el}	Electrical efficiency
Eff_{th}	Thermal efficiency
Eff_{SOFC}	SOFC efficiency
Eff_{tot}	Total (CHP) efficiency
FTU	Turbine-driven fan unit or fan turbine unit
FU	Fuel utilisation
GCC	Grand composite curve
GCU	Gas cleaning units
GT	Gas turbine
HEN	Heat exchanger network
HEX	Heat exchanger
ir	Interest rate
KPI	Key performance indicator
LHV	Lower heating value (kJ/kg)
LPG	Liquefied petroleum gas
m	mass flowrate (kg/s)
MER	Maximum energy recovery
MO	Multi-objective
NPV	Net present value (MEUR)
OPEX	Operative expenditure (MEUR/yr or EUR/MWh)
PFD	Process flow diagram
P_{comp}	Power consumed by the compressors/fans of the system (kW)
P_{prod}	Gross power produced by the SOFC LSM (kW)
P_{pump}	Power consumed by the pumps of the system (kW)
P&ID	Pipping and instrumentation diagram
Q_{cu}	Heat that is absorbed by the cold utility (kW)
Q_{hw}	Heat used to produce hot water at 1.0.1325 bar and 65 °C(kW)
Q_{turb}	Thermal power consumed by the turbine of the FTU (kW)
RR	Recirculation ratio
SOFC LSM	Solid oxide fuel cell large stack module
STB	Steam to biomass ratio
T	Temperature (°C)
T_{in}	Stream inlet temperature before a heat exchange (°C)
T_{out}	Stream outlet temperature after a heat exchange (°C)
ΔT_{min}	Minimum temperature difference between a hot and cold stream (°C)

1 EXECUTIVE SUMMARY

BLAZE (Biomass Low cost Advanced Zero Emission small-to-medium scale integrated gasifier-fuel cell combined heat and power plant) aims at developing an innovative highly efficient and fuel flexible small and medium-scale biomass CHP (Combined Heat and Power generation) technology. The purpose of the current deliverable is to present a range of optimum BLAZE plant layouts based on the outcomes of a techno-economic optimisation and assessment. The BLAZE plant uses a double bubbling fluidised bed gasifier (DBFBG) and an SOFC large stack module (LSM) as main technologies. It gasifies biomass waste to produce electricity (at a scale of 25 kW_e) and steam. The reported results summarise the most convenient operating conditions of the selected variables and provide an optimised heat exchanger network (HEN). The applied methodology follows a systematic procedure for multi-objective (MO) (i.e. maximum electrical efficiency, maximum thermal efficiency and minimum heat exchanger –HEX area) optimisation, and the solving strategy combines process flow modelling in steady state and mathematical programming. The selected layouts are further analysed from the economic point of view (net present value – NPV) to understand the conditions that could make the BLAZE plant financially attractive.

The starting point of the current report is the work developed in D4.1. That work, together with an analysis of the anode off-gas (AOG) treatment options (described in D4.3), derived into a scenarios analysis (Section 5) that set the basis of the layout to be optimised for the design of the BLAZE plant. The optimisation approach and results are described in Sections 6 and 7. The obtained outcomes in the current deliverable are used as reference in T4.4 (design of the turbine-driven fan unit or fan turbine unit –FTU) and in T4.6, which proposes the pilot plant process flow diagram (PFD) and piping and instrumentation diagram (P&ID). The obtained results will also be used in WP5 and 6. The current deliverable corresponds to T4.2 and T4.5.

The results point out that the combined efficiency of the plant can go higher than 80 %. At its turn, the electrical efficiency can go up to 49 %.

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2 INTRODUCTION

The BLAZE plant uses a DBFBG and a SOFC LSM as main technologies. The syngas, before being used in the SOFC LSM, has to be appropriately cleaned to avoid the contamination of the electrochemical device. The selected gas cleaning units (GCU) in BLAZE plant work at high temperature.

The BLAZE pilot plant uses an existing 100 kW_{th} DBFBG in combination with an SOFC LSM of 25 kW_e. The combined heat and power generation (CHP) capacity of the plant is an important characteristic, since BLAZE plant design considers not only the electricity generated as product, but also the possible integration with specific heating demands of external units.

The different technologies that compose the BLAZE plant are:

- Biomass waste handling and pre-treatment (not modelled)
- The DBFBG with inserted filter candles
- Hot GCU (removal of H2S, HCl, and alkali compounds) and tar reformer
- SOFC LSM
- AOG recirculation device (the FTU)

These technologies are modelled in the current deliverable by using key design variables that will be optimised, following a conceptual design approach. First, different recirculation points are evaluated in the scenarios analysis of Section 5. Then, the selected layout to be optimised is systematically evaluated in the MO optimisation approach, pointing out the trade-offs among electrical efficiency, thermal efficiency and cost (represented by the total needed HEX area). The desired project targets are; <u>50 % electrical efficiency, 40 % thermal efficiency, total capital expenditure (CAPEX) < 4000 EUR/kW, operative expenditure (OPEX) < 0.05 EUR/kWh and electricity price below 0.10 EUR/kWh</u>. The obtained results point out the main influencing variables, the optimum operating conditions, the overall system performance for the selected biomass waste (hazelnut shell), the HEN, for selected optimum configurations, and the market competitiveness of the BLAZE plant.

2.1 Objectives and scope of the document

The present document corresponds to T4.2 “Process modelling and validation” and T4.5 “Techno-economic optimization of conceptual CHP plant designs”. Both tasks end at M36, aiming at including the results of the tests in the modelled units (WP2 – Gasifier tests, WP3 – SOFC LSM tests and WP6 – overall plant tests), and at contributing to the economic and environmental life cycle assessments with the detailed inventory of streams characteristics. The current deliverable focuses on the conceptual design of the BLAZE plant; the plant model, optimisation and analysis are used to depict the most favourable

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working conditions and layout of it. These results are extrapolated to the pilot plant design in D4.3 and D4.5.

The document is structured as follows: Section 3 summarises the followed methodology. The BLAZE plant layout and model are described in Section 4. Section 5 mainly includes the scenarios analysis, that constitute the basis for the plant layout definition. Section 6 focuses on the optimisation results of the layout defined during the scenarios analysis.

3 METHODOLOGY

System modelling, optimisation and sensitivity analysis are used to evaluate different BLAZE scenarios. The methodology can be divided into three main steps:

1. **Scenarios evaluation**. This first step includes the analysis performed in D4.1 and the investigation of subsequent scenarios derived from these very first calculations. Overall, the considered scenarios take into account the three possible recirculation points, i.e. (i) AOG to the SOFC LSM anode inlet, (ii) AOG to the gasifier combustor and (iii) AOG to the gasification chamber, the use of an AOG gas turbine (GT) and the possibility of not taking advantage of the AOG calorific value. The evaluation of scenarios also considers different plant pressure management strategies, i.e. pressurised gasifier vs suction blower after the GCU.
2. **MO optimisation**. From the previous step, one plant layout is selected for optimisation. The results of the optimisation indicate the advised operating conditions from a list of decision variables, and the HEN of the BLAZE plant.
3. **Economic sensitivity analysis**. Once selected the BLAZE plant operating conditions and HEN, the economic sensitivity analysis aims at finding the variables (selected prices and costs) with the most influence on the NPV of the BLAZE plant and the conditions under which it becomes positive.

Process simulation is the basis for the scenarios evaluation and for the process optimisation (models performed in Aspen Plus V10). Process optimisation includes process simulation and process heat management, to elucidate the most suitable operating conditions and HEN for the BLAZE plant.

3.1 Simulation

After a preliminary analysis of the targeted process, including the compilation of the operating windows of the BLAZE plant units (see Section 6 of D4.1) and a preliminary analysis of scenarios (see Section 8.6 of D4.1), in the current deliverable we model and evaluate the BLAZE plant scenarios corresponding to three different recirculation points, the use of a GT to take advantage of the AOG calorific value, and the direct vent of the AOG, without a pre-determined HEN. In the scenarios evaluation, heat integration is assessed by the calculation of the problem table.

In the optimisation, heat integration is represented by a mathematical model in AMPL (A Mathematical Programming Language, a specific modelling environment for the formulation and solution of mathematical programming models). The adapted "SYNHEAT temperature-stage" HEN superstructure of Yee and Grossmann [1] in Martelli and co-workers [2] is used in the optimisation. The results indicate the most efficient combination of cold and hot streams by minimising the number of connections. The results of the scenarios evaluation, i.e. a preferred layout and the essential trade-offs among the selected performance indicators, are the starting point of the optimisation.

3.2 Performance indicators

The performance indicators were defined in D4.1 and are summarised below.

The cold gas efficiency (CGE) calculates the performance of the gasifier by considering the lower heating value (LHV) of the involved streams.

$$CGE = \frac{\dot{m}_{syngas,ar} \cdot LHV_{syngas,ar}}{\dot{m}_{biomass,ar} . LHV_{biomass,ar} + \dot{m}_{LPGs,ar} . LHV_{LPG,ar} + \dot{m}_{AOG,ar} . LHV_{AOG,ar}} \quad (1)$$

Where, $\dot{m}$ is the mass flowrates in kg/s, the LHV is expressed in kJ/kg; *ar* refers to "as received" (including moisture content and ashes, when present).

The SOFC efficiency (Eff_{SOFC}) takes into account the gross SOFC LSM power produced (P_{prod} in kW), divided by the inlet fuel, that is the syngas after the GCU (when available, after AOG mixing).

$$Eff_{SOFC} = \frac{P_{prod}}{\dot{m}_{syngas,ar} . LHV_{syngas,ar}} \quad (2)$$

The electrical efficiency (Eff_{el}) considers the net power produced in the system (that is, the gross power from the SOFC LSM P_{prod} minus the power consumed by system' compressors and pumps, in kW) divided by the total inlet calorific value into the system, provided by the biomass and LPG streams.

$$Eff_{el} = \frac{P_{prod} - (\sum P_{comp} + \sum P_{pump})}{\dot{m}_{biomass,ar} . LHV_{biomass,ar} + \dot{m}_{LPGs,ar} . LHV_{LPG,ar}} \quad (3)$$

The thermal efficiency (Eff_{th}) considers the total amount of heat used to produce hot water at 1.01325 bar and 65 °C (Q_{hw} in kW) using the COG stream (from 140 °C to 50 °C), the total amount of heat that is absorbed by the cold utility (Q_{cu} in kW), which in our case is steam produced at 5 bar and 220 °C (i.e. the preliminary agreed steam conditions needed at the turbine of the FTU), and the thermal power consumed by the turbine of the FTU (Q_{turb} in kW).

$$Eff_{th} = \frac{Q_{hw} + Q_{cu} - Q_{turb}}{\dot{m}_{biomass,ar} . LHV_{biomass,ar} + \dot{m}_{LPGs,ar} . LHV_{LPG,ar}} \quad (4)$$

The CHP performance, or total efficiency (Eff_{tot}), is the sum of Eff_{el} and Eff_{th}. One of the purposes of our plant design is to decrease as much as possible the consumption of the fossil fuel. The selected objective functions used in the optimisation are Eff_{el}, Eff_{th} and HEN area.

3.3 Optimisation

The present work adapts the multi-period approach described in Pérez-Fortes et al. [3] to consider one unique period, and the queuing MO optimisation algorithm QMOO is replaced by ev-MOGA, an elitist MO evolutionary algorithm which is available via the MATLAB central file exchange [4]. The sequential steps of the MO optimisation methodology consist of (i) the calculation of the mass and energy balances, (ii) the system energy integration and (iii) the evaluation of the performance indicators.

As a result of the optimisation, the Pareto front will allow the decision-maker to visualise the consequences of the selected choice. The selected objectives, maximisation of Eff_{el} and maximisation of Eff_{th} are conflictive among them, as in general, one increases if the other decreases (i.e. ideally, $Eff_{th} = 1 - Eff_{el}$). The third objective, the HEN area, acts as an economic criterion (capital cost), as the size of the plant is fixed. In Section 6.2, the selected non-dominated solutions reported correspond to the extremes of the Pareto frontier and to a weighted distance solution (Euclidean distance to the utopian point – 0.4 of Eff_{th} and 0.5 of Eff_{el}, and the minimum HEN area found during the optimisation). The aim of the optimisation is twofold: (i) to optimise the process design specifications, and (ii) to propose a HEN structure (see next section).

3.3.1 *Heat integration*

The SYNHEAT superstructure (see Figure 1) is used to determine the HEN layout. It stands for the type of streams connections and overall outline, originally proposed in [5]. The proposed representation stands for a "stage-wise" superstructure which allows for different streams matching; within each stage, potential exchanges between hot and cold streams can happen.

The BLAZE plant results in a threshold problem; i.e. only requiring a cold utility (thus, it produces enough heat for its process, and still has excess of it). The calculation of Eff_{th} takes into account that all the system heat that needs to be released via the cold utility is used to produce steam at 5 bar and 220 °C (i.e. the steam conditions needed at the turbine of the FTU). In the scenarios evaluation and in the optimisation it differs in:

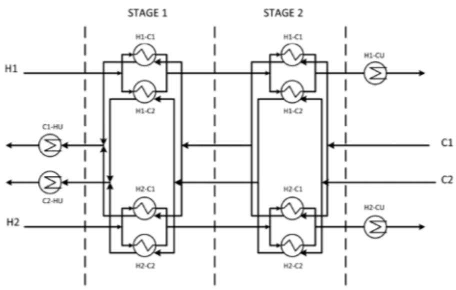

Figure 1: The SYNHEAT superstructure proposed in [1], from [6].

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- In the first, the FTU is modelled and we take into account its ad-hoc steam needs and generation for every considered layout.
- In the second, with the aim of being as flexible as possible with the BLAZE plant possibilities, the FTU is not modelled for the optimisation, but we take into account the overall amount of steam produced. The specific turbine steam needs have been determined for only the selected Pareto cases. We modelled and evaluated the use of a commercial blower to recirculate the AOG towards the combustor. This device was included in the model used for optimisation; with its power consumption we are considering the impact of the flow and temperature of the AOG recirculation towards the combustor in the Eff_{el}.

3.4 Economic sensitivity analysis

The modelling task provides with the needed input for the economic analysis of the BLAZE plant. We use in this section the net present value (NPV) as the selected metric for the economic evaluation of the project. We follow the methodology described in [7]. The competitiveness of the BLAZE plant (i.e. NPV zero and above) is analysed via the sensitivity analysis of selected prices and costs. This analysis shows which variables have the largest influence on the NPV. The most important variable in our case is the price of the electricity.

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4 DESCRIPTION OF THE MODEL

4.1 Description of the process

The system has two main units: the fluidised bed gasifier, that converts biomass waste into syngas and the SOFC LSM, that converts the syngas into power. The main places where the AOG can be recirculated, are three: (i) the SOFC LSM anode inlet, (ii) the gasifier combustor and (iii) the gasification chamber. The hot GCU, including chlorine and sulphur compounds separators and tars reformer, together with the catalytic filter candles at the outlet of the gasification chamber, are crucial to keep the levels of slow tars, fast tars, sulphur compounds, halogen and alkali compounds low, to avoid carbon deposition, fouling and corrosion, specifically as needed by the SOFC LSM. The heat of the gasification process, in the current pilot gasifier, is provided by LPG and residual char from gasification. One of the purposes of the current deliverable is to elucidate the conditions under which is amount of LPG can be minimised, so as to produce electricity without fossil fuels.

Biomass waste (specifically, hazelnut shell) and steam are fed to the gasifier. The gasifier is a dual fluidised bed gasifier or indirectly heated steam gasifier, where steam gasification is separated from combustion. The unit includes a ceramic filter candles filled with commercial Ni-catalyst pellets (thus, hot gas cleaning) particles removal and decomposition of tar and ammonia [8]. A flue gas is produced in the combustor and acts as a heat source in the plant. The syngas stream moves towards the GCU, after cooling, where different units intervene to remove Cl compounds, S compounds and tars. The clean syngas is then preheated to the required SOFC LSM inlet temperature. The air supply to the gasifier and the fuel cell is controlled by two blowers; both streams are preheated to the desired gasification and fuel cell temperatures. The COG, before being released, is used to produce hot water.

The gasifier and the SOFC LSM have to accomplish with specific pressure conditions. In order to use a pressurised gasifier in the BLAZE plant, an ad-hoc screw feeder should be used, which has to be appropriately designed to avoid inner hot gas to flow back in the feeder, pyrolysing inlet biomass. In the SOFC LSM, the anode pressure should always be above the cathode pressure, in a range of 5 – 30 mbar. Ideally, it should operate above atmospheric pressure. The maximum absolute pressure that the cathode can tolerate is 1.09 bar. Pressure management in the overall plant affects the design of the FTU, as the recirculation pressure can change and depends on the recirculation point. Tolerable pressure losses and plant constraints were summarised in D4.1.

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4.2 Process modelling

The mass and energy balances are performed in Aspen Plus V10 software. The following tables summarise the different scenarios that have been evaluated along the BLAZE project. The purpose was to elucidate the most convenient plant layout for optimisation, and very important of our project, the most convenient location of the AOG recirculation point. The first modelled plant layouts correspond to D4.1. Among the different options analysed, three configurations were selected for further analysis: these are summarised in Table 1. In these three scenarios, the HEN was fixed and it was proposed based on the most probable heat exchange that would occur in the plant, according to experimental experience. Within these first plant models, steam generation was modelled (production at 400 °C and 1 bar, according to the gasifier needs), and the biomass consumption was fixed, based on preliminary calculations of the syngas needed to obtain the desired power (i.e. 20 kg/h of syngas). The results of these three simulations are reported in Section 5.1. Next, Table 2 summarises two more cases that were analysed to interpret the main differences between the use of the pressurised gasifier versus the use of a suction blower downstream the tar reformer. Further sensitivity analysis of selected variables, revealed the importance of the AOG to decrease the consumption of LPG in the gasifier combustor, as indicated in Section 5.2. Also in this section, there is a brief summary of the calculations performed to analyse the role of the FTU recirculating AOG towards the anode inlet; this is further described in D4.3.

Table 1. Cases (different plant layouts or operating scenarios) analysed in D4.1 (see Section 5.1). The HEN is fixed.

Name	Description
Case B	Pressurised gasifier. AOG sent to the SOFC LSM inlet stream (RR=0.5) and the rest to the gasifier combustor. The FTU is used in the AOG to LSM stream. FU global = 0.75.
Case D	Pressurised gasifier. AOG sent to the gasifier combustor (without FTU). Case D1 (FU = 0.6) and D2 (FU = 0.75). Without FTU.
Case F	Analogous to B, without a pressurised gasifier but with a suction blower after the tar reformer. FU global = 0.75.

Table 2. Cases analysed in Section 5.2. The HEN is not fixed.

Name	Description
Case D	Pressurised gasifier. AOG sent to the gasifier combustor without FTU. FU = 0.75. Cases D50 and D200, with or without AOG steam condensation (recirculation at 50˚C or 200˚C).
Case F	Cases F1, FU = 0.75 and AOG use in the combustor, at high T, and F2, FU = 0.75 and AOG combustion after water separation. Suction blower after tar reformer.

Finally, Table 3 summarises the selected cases for further simulation and comparison, so as to conclude the final layout to be optimised. In these cases, biomass feedstock flow was adapted to produce, in each particular case, 25 kW_e gross in the SOFC LSM.

Table 3. Cases analysed in Section 5.3.

Name	Description
Case 1	Base case; BLAZE plant without AOG use.
Case 2	AOG recirculation to the gasification chamber.
Case 3	AOG recirculation to the SOFC LSM anode inlet.
Case 4	AOG recirculation to the gasifier combustor without FTU.
Case 5	AOG recirculation to the gasifier combustor with FTU.
Case 6	AOG used in a GT.

4.2.1 Fluidised bed gasifier

We modelled the gasification chamber separated from the combustion chamber, to take into account the heating needs (i.e. fuel needed) to have a balanced gasifier [9,10]. It is assumed that the gasifier works at 1.29325 bar, is isothermal (per separate, gasifier chamber and combustion chamber), char is 100 % carbon and that the overall heat needed in the gasification process is provided by the gasifier combustor. The feeding system is not modelled. The conversion of N, S and Cl into the contaminants NH_3, H_2S and HCl is assumed to be complete.

The inlet streams to the gasifier are biomass waste, steam (to the gasification chamber), air, LPG (modelled as propane) and AOG in Cases 2-5 (to the combustion chamber). Internally, also a fraction of char, from the gasification process, is burnt into the combustor.

4.2.2 GCU

The gas cleaning units in the BLAZE plant include [11]: the in-bed gas cleaning by a calcined dolomite bed, the catalytic filter candles, the alkali-based sorbent reactor that separates Cl compounds, the sorbent reactor that separates S compounds, and the tar reformer.

The syngas cleaning units are modelled as:

- An RStoic reactor simulating the catalytic filter candles, where methane, toluene, benzene and naphthalene react with water to produce CO and H_2. These reactions are considered to take place at a temperature that is 70 °C lower than the gasification temperature.
- Two heat exchangers that adapt the temperature to 400 °C and 550 °C; the two selected operating temperatures for S and Cl separation, and for the tar reforming, respectively.
- The HCl adsorber, H_2S adsorber and tar reformer are simply modelled as a component separator that splits all the contaminants from the syngas before the SOFC LSM.

4.2.3 SOFC LSM

The SOFC LSM model compiles the modelling approaches in [3,12–14]. The anode inlet pressure is 1.07 bar and the cathode inlet pressure is 1.06 bar. The model is 0D and considers that the inlet gases are heated to a temperature of 700 °C, the outlet temperature (AOG and COG) is 790 °C, and that the electrochemical and chemical reactions occur at an average reactor temperature (calculated as an average of the inlet and outlet temperature (745 °C with base conditions)). The model consists of an anode block modelled by two RGibbs. The second RGibbs receives the O_2 from the cathode block, simulated as a component separator that splits the O_2 required for the electrochemical reaction. The inlet amount of air is controlled to reach the desired outlet temperature.

4.2.4 FTU

This unit is modelled using the compressor and turbine units from Aspen Plus. The connecting condition implies that the steam turbine has to provide all the power needed by the fan. It depends on the flowrate and the composition of the recirculated stream, as well as on the temperature of the AOG. The analysis performed considers that the AOG temperature at the turbo-fan inlet is in the range of 40-50 °C, which may be challenging because of operational limitations. It is assumed that the turbine has an isentropic efficiency of 40% (mechanical efficiency of 100%), and the inlet steam is at 5 bar and 220 °C. The discharge pressure is 2.5 bar. The turbo-fan is assumed to work with an isentropic efficiency of 60% (mechanical efficiency of 100%).

4.2.5 Compressors, heaters and coolers

The components such as pumps for water supply, blowers for air supply and gas circulation, heaters and coolers, are modelled using standard Aspen Plus library components. The performance of the blowers is determined based on the isentropic efficiencies (60%, and mechanical efficiencies of 100%). The pump efficiency is taken to 80%. For the design of the HEN, a minimum temperature difference between the hot and cold streams (ΔT_{min}) of 30 °C is assumed.

4.2.6 GT

We have also considered the burning of the AOG and its use in a GT. All the AOG is sent to a downstream combustor. The burner has a stoichiometric combustion. The burner has three inlet streams: AOG, air and steam. The AOG enters the unit at 790 °C and 1.04 bar. Air enters at the same pressure and temperature, while steam enters at atmospheric pressure. The inlet amount of steam corresponds to an excess ratio of 3 [15]. The burner is adiabatic and the outlet temperature results in 1176 °C. The flue gas is vented at 140 °C.

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5 PRELIMINARY FLOWSHEET AND PILOT PLANT DESIGN STEPS

The following sections summarise the evaluated BLAZE plant layouts and describe the results obtained, that drive us to consider one specific layout for optimisation.

5.1 Results from D4.1

In D4.1 three configurations were selected for analysis; these were summarised in Table 1. In these three scenarios, the HEN was fixed. Within these first plant models, steam generation was modelled, and the biomass consumption was set to produce 20 kg/h of syngas. Recirculation was present in Cases B and F, sending half of the AOG flow to the gasifier combustor and half to the SOFC LSM anode inlet. In Case D, all AOG stream was sent to the gasifier combustor, via a conventional compressor. Two cases D were taken into account, with different FU value (D1, where the global FU value is 0.6, and D2, where the global FU is 0.75). Note that AOG steam condensation was only considered in these results if the FTU was present.

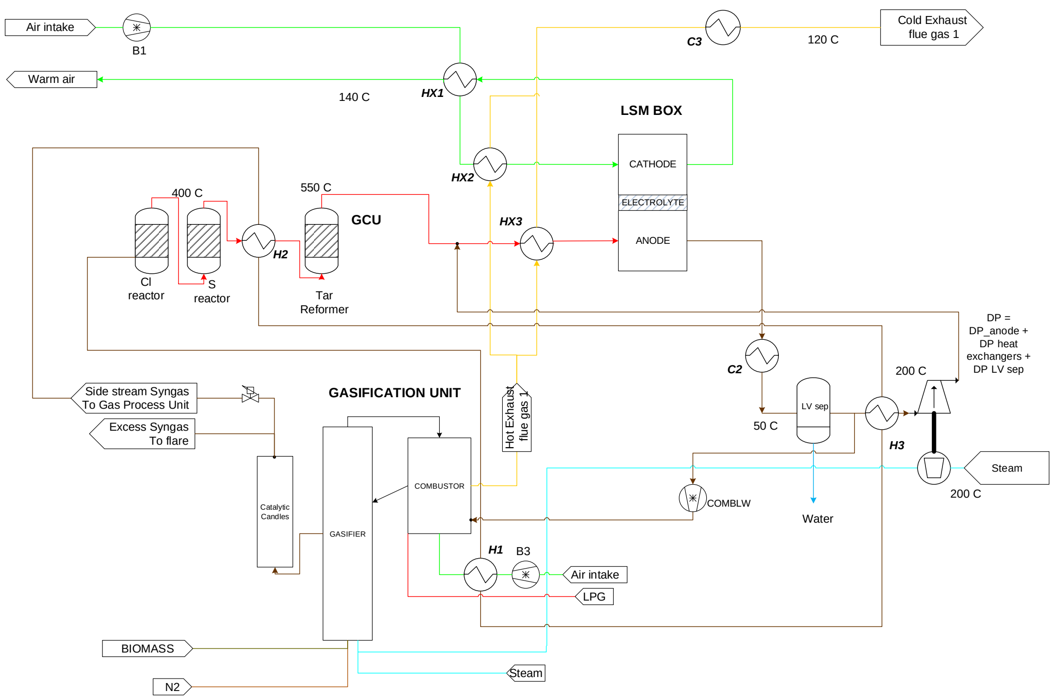

Figure 2. Case B PFD of the BLAZE plant. The AOG is sent to the SOFC LSM inlet stream (RR = 0.5) and the rest is burnt in the gasifier combustor, with specific heat exchange (same configuration for Cases D – all the AOG is sent to the combustor and the FTU unit is not used, no condensation foreseen- and F; Case F has moreover a cooler, C4, downstream the tar reformer and before the downstream compressor).

Table 4 summarises the main results obtained for the different scenarios.

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Table 4. Summary of main results (D4.1).

Results	Case B	Case D1	Case D2	Case F
Power SOFC (kW)	26.95	22.40	27.00	26.95
Wnet (kW)	24.55	20.43	24.69	24.62
Syngas LHV (ar) (MJ/kg)	12.47	12.47	12.47	12.47
Syngas flow (kg/h)	15.9	15.9	15.9	15.9
Syngas composition (mole fraction)				
H_2	0.49	0.49	0.49	0.49
CO	0.25	0.25	0.25	0.25
CO_2	0.11	0.11	0.11	0.11
H_2O	0.15	0.15	0.15	0.15
CH_4	0.004	0.004	0.004	0.004
In biomass (kW)	58.6	58.6	58.6	58.6
CGE	0.65	0.64	0.62	0.64
Eff_{SOFC}	0.49	0.41	0.49	0.49
Eff_{el}	0.34	0.32	0.33	0.34

Concluded in D4.1: The combined heat and power efficiency could go up to 70 %, within selected working conditions and specific layouts. The importance of selecting the most favourable operating conditions and layout for the (i) gasifier and the (ii) SOFC LSM individually was pointed out, as the system evaluation of different plant layouts showed similar final results (i.e. similar efficiency values) (Table 4). It was therefore pointed out that there was place for further analysis of (i) the role of the AOG recirculation towards the SOFC LSM anode inlet (identification of SOFC performance advantages, if any); (ii) the role of the AOG recirculation towards the gasification chamber of the gasifier (the same, identification of gasification performance advantages, if any); and (iii) the role of the AOG recirculation towards the gasifier combustor (importance of AOG composition and temperature). The next scenarios specifically focus on these.

5.2 Scenarios pre-evaluation

Before analysing the role of the recirculation in the different layout possibilities, in the current section we (pre-)evaluate the impact of the AOG on the global system efficiency, via its impact on the gasifier combustor, and we detail the FTU treatments options when the AOG is recirculated to the anode inlet. A FU=0.75 is therefore selected as the base value in the SOFC LSM to produce the required power.

First, from the results obtained in the previous section, we evaluated the impact of the AOG temperature recirculation, in case D, with FU=0.75. To remember from D4.1 that there was an interest in steam condensation before anode inlet recirculation; thus, the AOG stream sent to the gasifier combustor had as well lower steam fraction in case B. Note that in the current evaluations the HEN is not predetermined, but heating and cooling needs are of concern. See in Figure 3 the layout of case D.

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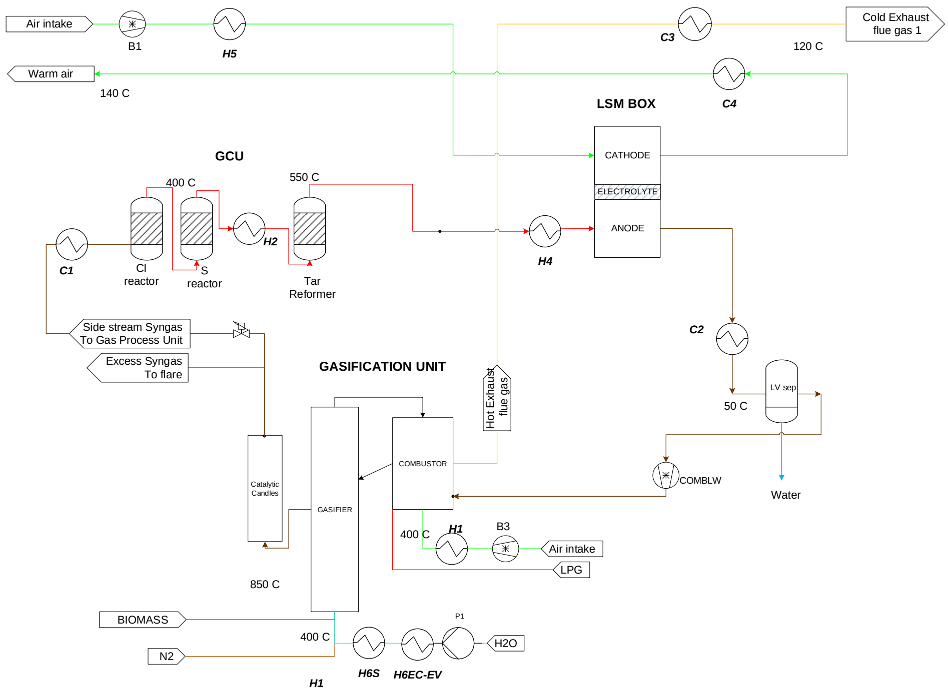


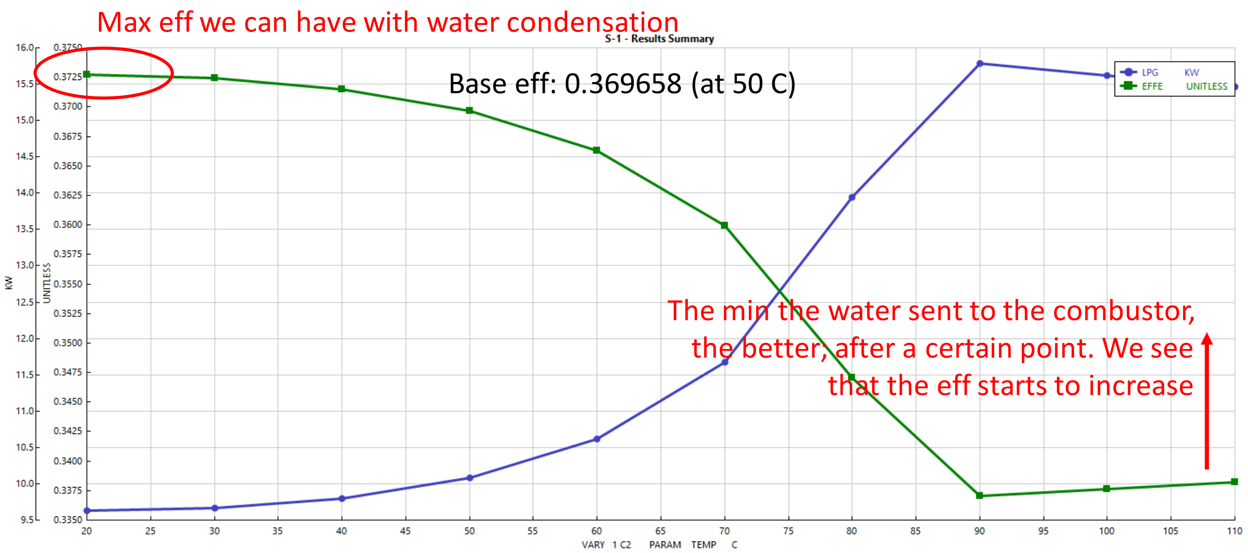
Figure 3. Case D PFD of the BLAZE plant. Through C2 we adapt the temperature of the AOG to 50 or 200 °C to study the effect of steam in the gasifier combustor.

The results in Table 5 point out the effect of AOG steam condensation in the electrical efficiency. We see that LPG need decreases, as the AOG temperature decreases, showing that the composition of AOG (water fraction) plays an important role. The enthalpy values have more impact in the final efficiency in this case, compared to a higher inlet temperature (but with higher water fraction). The efficiency values of Case B and D50 are therefore very similar.

Table 5. Comparison of Cases B, D200 and D50.

Results	**B**	**D200**	**D50**
Power SOFC (kW)	26.95	27.00	27.00
Wnet (kW)	25.43	24.99	25.40
Syngas LHV (ar) (MJ/kg)	12.45	12.45	12.45
In biomass (kW)	58.64	58.64	58.64
Q_{in} LPG (kW)	9.34	14.17	10.08
T AOG to combustor (°C)	80.00	232.00	80.00
CGE	0.68	0.64	0.68
Eff_{SOFC}	0.49	0.49	0.49
Eff_{el}	0.37	0.34	0.37

These results referred us to perform a sensitivity analysis to study the role of AOG and air inlet combustion temperatures. The reported behaviour in Figure 4 points out the dual benefit of the AOG temperature customisation before the combustion: steam condensation and AOG heat up, to move towards the minimisation of the consumption of LPG. Therefore, it is deduced that, after condensing water, if the AOG is further heated up, the amount of LPG consumed can be further decreased.



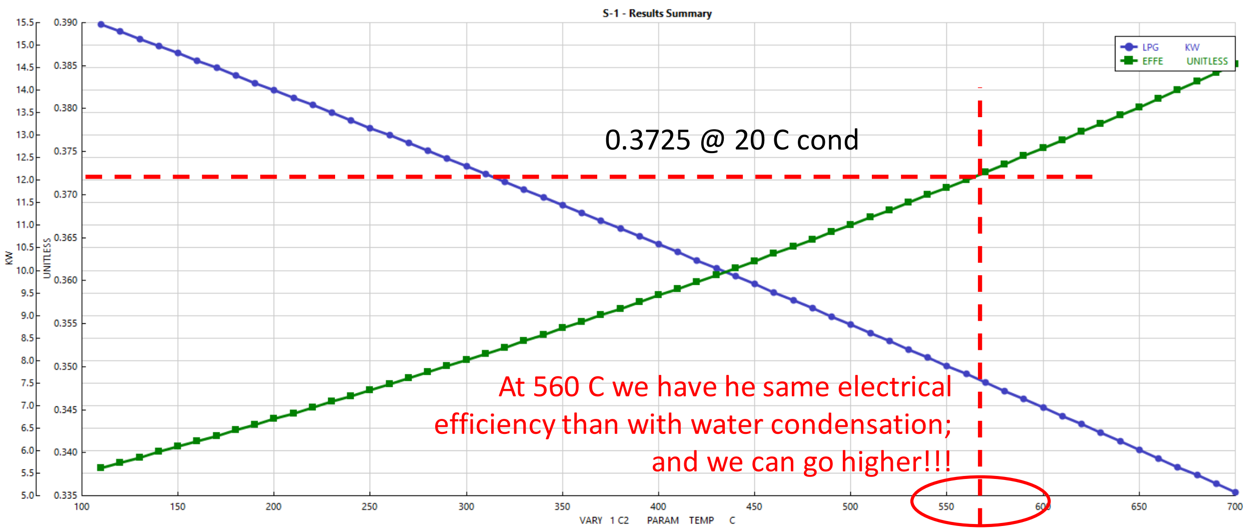


Figure 4. Change of AOG steam condensation temperature (C2) using Case D50 as starting point. Effects on the amount of kW of LPG consumed and on the Eff_{el}. Top graph, in a C2 temperature range 20-110 °C, in the bottom graph, in a range that goes up to 700 °C.

The sensitivity analysis results in Figure 5 indicate that the air temperature also influences the amount of LPG consumed in the gasifier to fulfil the gasification needs. Eff_{el} can change from 0.37 up to 0.39 with higher inlet air temperature. Thus, this temperature, together with the AOG temperature, are identified as important influencing variables in the BLAZE plant.

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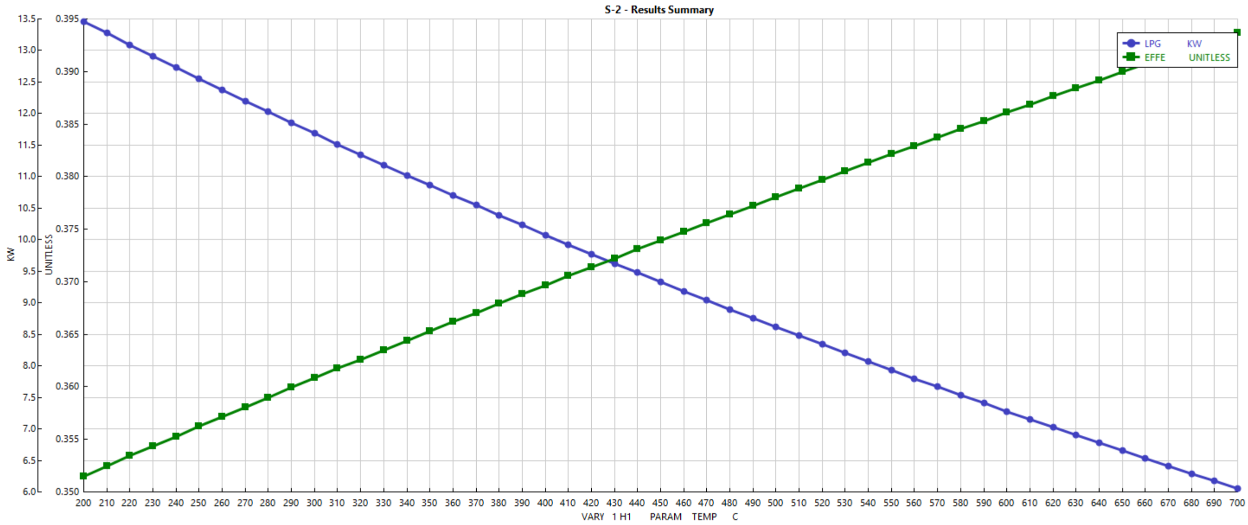

Figure 5. Change of combustion air temperature (H1), with Case D50 as base case.

In an analogous way, the role of AOG and air temperatures were studied in Case F. Figure 6 represents it, having as main change a compressor downstream the tar reformer (gasifier working at atmospheric pressure).

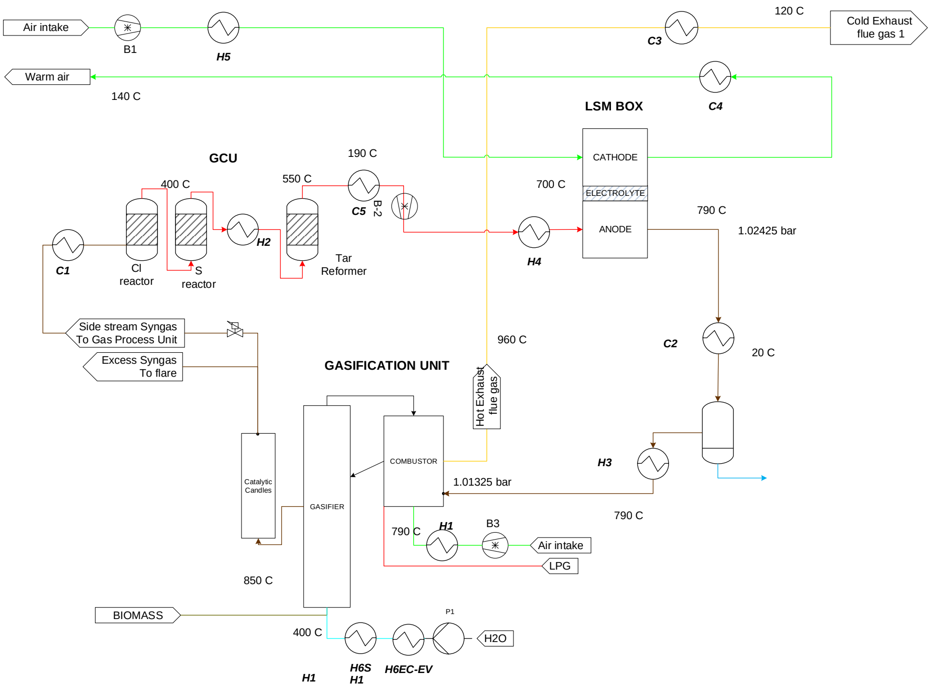

Figure 6. Case F2 PFD of the BLAZE plant. Case F1 does not consider condensation.

In Figure 7 the sensitivity analysis for Case F is represented when, following the previous results, the inlet air temperature is fixed at a maximum temperature of 790 °C. Moreover, AOG steam condensation is fixed

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at 20 °C, and the heat exchanger called H3 adjusts the AOG combustor inlet temperature. It is seen that, by increasing this last temperature, there is a temperature after which, the LPG consumption is zero. It is seen in Figure 8 that if this temperature continues to increase, the heat produced in the combustor is larger than the heat needed by the gasification process, thus, producing extra heat which is not consumed.

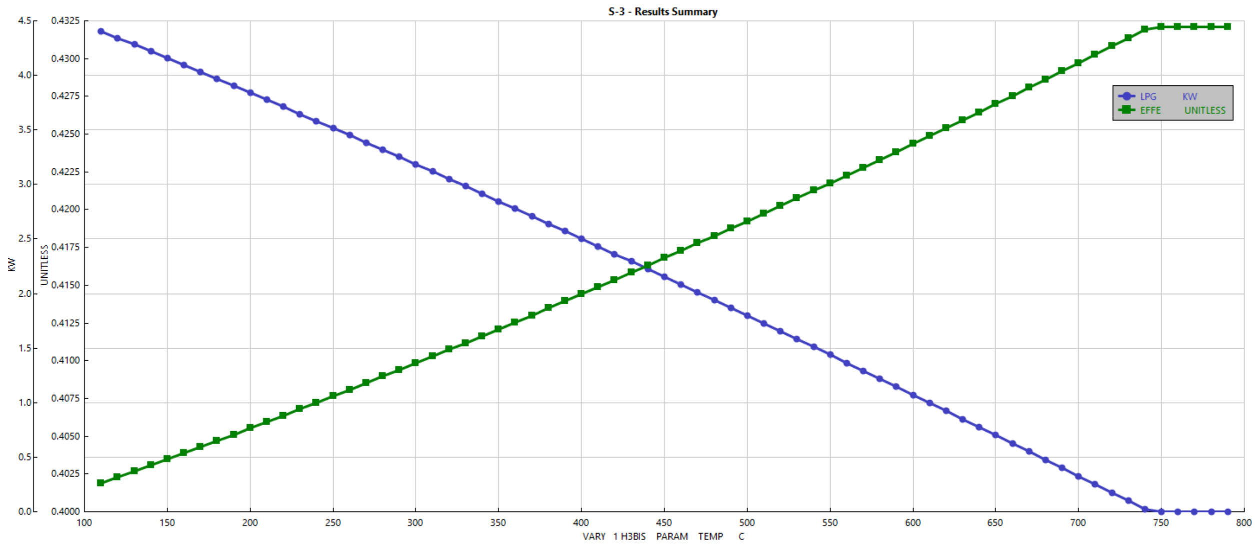

Figure 7. Change of AOG combustor inlet temperature (H3) using Case F as starting point. Effects on the amount of kW of LPG consumed and on the Eff_{el}.

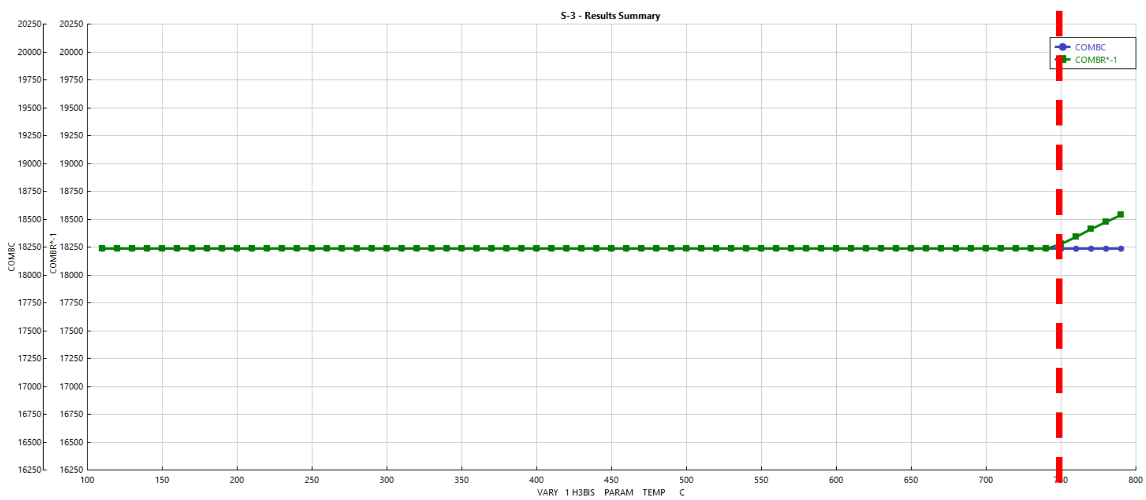

Figure 8. Change of AOG combustor inlet temperature (H3) using Case F as starting point. Effects on the heat consumed by the gasification process (in blue) and heat produced by the combustion chamber (in green), in W.

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Table 6 summarises the main results for cases D and F: case D considers the steam condensation and the maximum assumed temperature for the combustor inlet streams; case F is further divided into cases F2 and F1, which account respectively the scenarios with and without steam condensation.

Note the high Eff_{el} values for all the cases, above 40 %. At approximately equal efficiency, Case D (pressurised gasifier) is the selected configuration to move forward in the evaluation of scenarios. The reason is that the inclusion of an intermediate fan, as in Case F, requires additional heat exchange prior to the SOFC inlet with syngas (which is to some extent contradictory as in BLAZE plant hot gas cleaning is used).

Table 6. Comparison of Cases F1 and F2 (without and with steam condensation) and Case D (with steam condensation), AOG and air combustor inlet temperatures of 790 °C in Case F1. Note that when steam is condensated (Case F2 and D) the AOG temperature does not need to be as high as 790 °C to reach zero LPG consumption.

Results	**Case F1**	**Case F2**	**Case D**
Power SOFC (kW)	27.00	27.00	27.00
Wnet (kW)	25.34	25.34	25.62
Syngas LHV (ar) (MJ/kg)	12.47	12.47	12.45
In biomass (kW)	58.64	58.64	58.64
Q_{in} LPG (kW)	0.8604	0.0022	0.0004
T in AOG combustor (°C)	790	743	730
CGE	0.76	0.77	0.77
Eff_{SOFC}	0.49	0.49	0.49
Eff_{el}	0.43	0.43	0.44

Summing up, there is a clear efficiency improvement when AOG and air combustor inlet temperatures are increased, linked with the decrease of LPG consumption. When a dry AOG is recycled at the combustor inlet (that is steam condensation implemented), almost zero LPG consumption can be reached even at lower inlet temperatures. Following the first law of thermodynamics, it is shown that the amount of LPG consumed depends on the inlet AOG flowrate and its enthalpy. We would like to highlight that the increase of the electrical efficiency by decreasing the amount of consumed LPG is a step leading towards the production of fully green electricity, thus of interest for BLAZE plant.

A more detailed look into the effect of the AOG recirculation towards the inlet anode, evolved into a specific analysis that studied the AOG treatment possibilities to improve the SOFC performance, in collaboration with EPFL LAMD. We analysed the effect of the AOG recirculation into the SOFC voltage via a discretised SOFC model in gPROMS (developed at EPFL GEM group). We saw that the AOG recirculation was positive in certain circumstances and from the system point of view: when reforming is needed, steam generation needs can be decreased, as the AOG is mainly composed by steam and CO_2 (with smaller fractions of H_2 and CO). However, in BLAZE plant, the syngas has a minimal (almost zero) fraction of methane. In this case, see the effect of the recirculation fraction in Figure 9. The operating voltage at RR=0 and FU local and global = 0.75 is 0.78 V. The operating voltage at FU local and global = 0.6 is higher. However, if the inlet gas compositions changes, as it is the case when using the recirculator, the potential decreases; i.e. at FU=0.6 and RR=0.5, the potential goes down to approx 0.75 V. Note that the voltage, with hypothetical total steam condensation (and steam condensation at 50 °C) can improve, but cannot equal the initial potential without RR (at higher FU). However, if CO_2 and steam can be completely removed, the situation changes, as the AOG is thus a stream of only H_2 and CO. In this way, the fuel is therefore improved, providing a potential, even higher than the initial one, without RR. Based on these results, in D4.1, the vapour-liquid separator was placed in the AOG stream. As shown in the current section, this unit has also proven beneficial when the AOG is recirculated towards the gasifier combustor; i.e. AOG, is a residual stream with certain calorific value that can be used to improve the overall efficiency of the plant. However, the gas, has to be appropriately conditioned for such improvement.

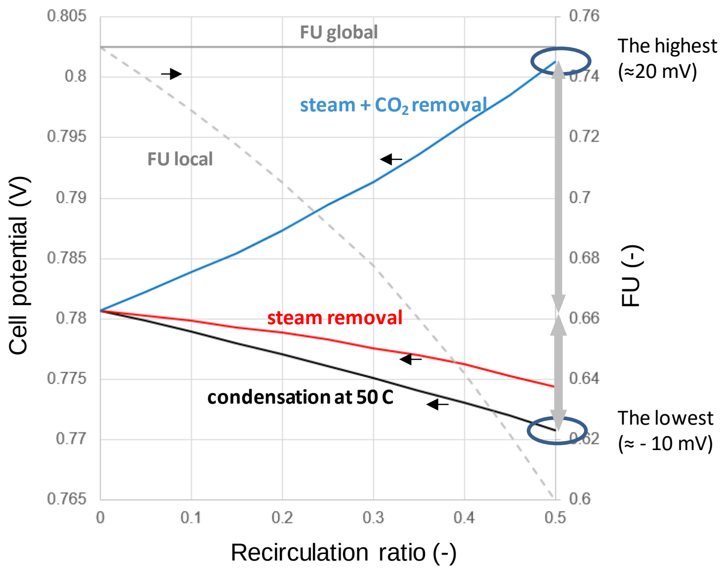

Figure 9. Change of the cell potential (left axis) as a function of the RR (x axis), at a constant global FU of 0.75 (SOFC model in gPROMS).

As summarised in D4.3, different treatment units have been therefore modelled to treat the AOG (compiled in Figure 10). The resulting operating voltages are collected in Table 7. Note that as previously mentioned, the best configuration counts with RR=0, or with CO_2 capture and steam condensation for improved operation. However, CO_2 separation at such a small scale is not yet commercially available.

Table 7. SOFC operating voltage at different AOG treatment conditions and layout configurations (with SOFC model in gPROMS).

Scenario	**SOFC potential (V)**	**SOFC electricity (kW)**
FU=0.6 RR=0.5 to SOFC LSM inlet	0.7483	24.138

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FU=0.75 RR=0	0.7747	24.990
AOG to gasifier (supply steam needs)	0.7492	24.166
Steam separation at 10°C, RR=0.5 (1)	0.7679	24.769
Methanation and steam reformer, RR=0.5 (2)	0.7483	24.136
Reversed WGS reactor, RR=0.5 (3)	0.7684	24.784
Hypothetical CO_2 separator, RR=0.5 (4)	0.7950	25.645

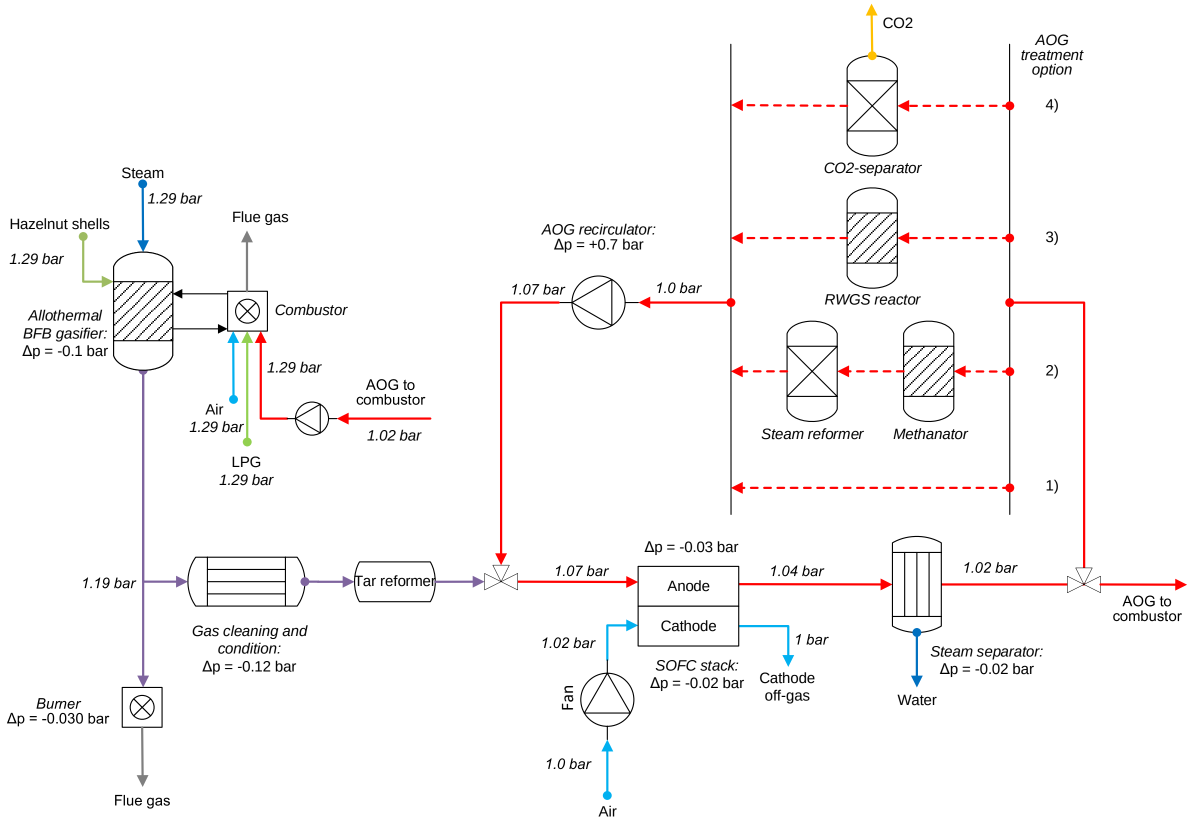

Figure 10. AOG treatment units considered in D4.3 (work developed together with EPFL LAMD).

The results provided in this section determine a base line for the scenarios in Section 5.3. Even though we have seen that the AOG recirculation towards the anode and gasifier inlets decrease the SOFC performance, we study in the following section the AOG temperature conditioning and recirculation points and their impact on the overall plant efficiency.

5.3 Scenarios evaluation

The BLAZE plant is a threshold problem; i.e. only cold utility is needed.

The results of the evaluation of the six cases are summarised in Table 8. The net electricity produced is around 24 kW_e in all cases, except when using the GT, the electricity produced goes up to 31.7 kW_e. The amount of LPG considerably decreases in Cases 3, 4 and particularly 5; when RR is 0.5 and in the last case, when the temperature of the AOG and air to gasifier combustor are increased. In order to calculate the

Eff_{el} value, we can see from the discussed numbers that the larger difference is marked by the remarkable decrease of LPG needs in Case 5 (compared to the biomass inlet decreased of Case 2, and the inlet LPG amounts in Cases 3 and 4).

The electrical efficiency is 10 percentage points larger in Case 5 than in Case 1, revealing the clear benefit of using the AOG within the plant. The electrical efficiency is 45 % in Case 6, compared to 44 % in Case 5, when the GT is used. The cooling water that can be generated via COG cooling down is also evaluated. Eff_{th}'s are higher in Cases 3, 4 and 6. The total efficiency is higher in Case 6, followed by Cases 3 and 4, Case 5 and Cases 2 and 1. However, as mentioned before, thermal efficiency will always be higher when more LPG is consumed. Therefore, as criterion for scenario selection for optimisation, we consider the Eff_{el}. Overall, we can see that the AOG use in the gasifier combustor decreases the use of LPG, more if inlet temperatures are increased. The recirculation of the AOG to the gasifier and to the SOFC results in more diluted syngas entering the SOFC LSM (thus, in a worst overall system performance). The results of Eff_{el} of Cases 5 and 6 are very similar. However, due to the added system complexity in Case 6 and the higher consumption of LPG, Case 5 is the selected one for optimisation. As conclusion of this section, Case 5 is the selected layout for optimisation.

Table 8. Modelling results for Cases 1 to 6. Global FU=0.75, STB=0.5, AOG steam condensation at 20 °C, recirculator FTU fan inlet at 200 °C, air to gasifier combustor at 400 °C, except for Case 5; where the recirculator FTU fan inlet temperature is given by the condensation temperature (20 °C) and the air to gasifier combustor temperature is 600 °C.

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Inlet biomass (kg/h)	10.29	9.15	10.25	10.29	10.29	10.29
Inlet air to combustor (kg/h)	28.56	35.62	31.01	32.06	22.23	28.62
Inlet LPG (kg/h)	1.26	1.32	0.62	0.67	0.11	1.26
Heat needed from the gasifier combustor (kW)	16.55	17.61	16.49	16.55	16.55	16.55
Inlet air to SOFC (kg/h)	381.59	429.31	351.68	381.59	381.59	381.59
Local FU	0.75	0.75	0.60	0.75	0.75	0.75
Global FU	0.75	0.75	0.75	0.75	0.75	0.75
STCR	0.39	0.34	0.21	0.40	0.40	0.39
AOG (kg/h)	23.22	35.51	37.34	23.23	23.23	23.22
Gross power SOFC (kW)	25	25	25	25	25	25
Gross power turbine (kW)						8.973
Net power (kW)	24.046	23.788	23.947	23.877	24.106	31.656
CGE	0.73	0.75	0.68	0.68	0.75	0.73
SOFC efficiency	0.50	0.47	0.50	0.50	0.50	0.50
Eff_{el}	0.34	0.37	0.39	0.38	0.44	0.45
Cooling water produced (kg/h)	189.68	213.61	174.52	189.68	189.68	189.68
Cold utility (kW)	6.23	8.06	18.75	18.35	4.72	17.41
Eff_{th}	0.22	0.28	0.45	0.45	0.25	0.38

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Total efficiency ($Eff_{el} + Eff_{th}$)	0.57	0.65	0.83	0.83	0.69	0.84
FTU						
ΔP (mbar)		250	60		270	
Steam needed (kg/h)		19.85	3.75		10.19/16.50	
Inlet fan T (°C)		200	200		20/200	
Total power needed from turbine (kW)		0.315	0.060		0.162/0.209	

[1]In Case 6, this value corresponds not to the AOG compressor, but to the GT air compressor and the GT flue gas compressor.

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6 OPTIMISATION RESULTS

The variables considered for optimisation, according to the results obtained from the previous section, are summarised in Table 9. The objective functions are Eff_{el}, Eff_{th} and HEX area. We considered forbidden matches among different HEXs, that correspond to prohibited hot-cold stream connections in the HEN design due to possible flammability issues. The forbidden matches selected are: (C4,H4), (C4,H2), (C1,H5), (C1,H1) and (C2,H1).

6.1 Decisions variables

Table 9 summarises the decision variables selected for the optimisation.

Table 9. Decision variables for optimisation.

Decision variable	**Range**
1. FU on the SOFC LSM (FU)	0.6 - 0.8
2. STB in the gasifier (STB)	0.33 - 0.98
3. Temperature of gasification (T Gasif)	750 - 850 °C
4. Fuel cell inlet temperature (T in SOFC)	690 - 750 °C
5. AOG cooling temperature (T C2)	20 - 300 °C
6. Temperature of inlet air to the gasifier combustor (T H1)	100 - 760 °C
7. Temperature of inlet steam to the gasifier (T H6S)	200 - 400 °C
8. Operating temperature of chlorites and sulphur compounds abatement units (T C1)	200 - 450 °C
9. Operating temperature of tar reformer (T H2)	550 - 700 °C
10. AOG inlet temperature to the gasifier combustor (T H3)	20 - 760 °C

6.2 Selection of optimum designs

We select the extremes of the Pareto front obtained from the optmisation (see in Table 10, columns Eff_{el}, Eff_{th} and HEN area). Moreover, we use the utopian point, as an ideal of the criteria values to find the closer solutions from the Pareto front. The utopian coordinates are Eff_{el}=0.5, Eff_{th}=0.4 and for the HEN area, the minimum area found in the whole range of executed scenarios was selected. See in the first column of Table 10 the data corresponding to the closest point to the utopian point.

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Table 10. Selected optimal process designs (extremes of Pareto front and closest point to utopian point) and performance.

Variable / criterion	Distance utopian	Eff_{el}	Eff_{th}	HEN area
FU	0.780	0.800	0.715	0.746
STB	0.333	0.330	0.967	0.330
Tgasif	782.475	751.173	837.502	839.452
TinSOFC	690.000	690.391	697.473	690.022
TC2 (AOG SOFC outlet)	28.705	25.873	26.207	186.869
TH1 (Air to combustor)	550.054	745.798	132.408	101.537
TH6S(superheated steam)	321.274	398.596	356.899	221.770
TC1 (syngas to GCU)	279.412	200.000	236.054	428.615
TH2 (GCU tar reformer inlet)	642.955	550.967	634.334	626.737
TH3 (AOG to combustor)	508.714	756.581	245.333	263.241
Eff_{el}	0.4547	0.4873	0.3443	0.3493
Eff_{th}*	0.3558	0.3052	0.4736	0.4093
Eff_{tot}	0.8105	0.7925	0.8179	0.7587
Heat exchanger area	9.980	11.543	13.614	6.727

The results in Table 10 reveal a total efficiency around 80 %. For the solution that is closer to the utopian point, the electrical efficiency is 45 %, the thermal efficiency is 36 % and the HEN area is 10 m^2.

The results presented are the extreme of the Pareto front and the utopian point, based on the reported weights above. However, other Pareto solutions from the Pareto front can be selected depending on the decision criteria of the decision maker; for instance, zero LPG consumption.

The HEN structure for each reported optimum case show an average number of 12 HEXs.

6.3 NPV sensitivity analysis

The current analysis is executed with the base case data of Case 5. Figure 11 depicts the total purchase costs share of the BLAZE plant. Among the overall investment needed, the SOFC LSM represents 46 % of it. It is followed by the gasifier and feeding system, HEXs and reactors and vessels.

summarise CAPEX and OPEX values of the BLAZE plant. Both of them are at least one order of magnitude higher than the targets.

Table 11. CAPEX and OPEX compared to the targets of BLAZE project.

	Calculated	Target
CAPEX (MEUR)	2.786	
CAPEX (EUR/kW)	111,440	< 4000
OPEX (MEUR/yr)	0.115	
OPEX (EUR/MWh)	534.5	< 50

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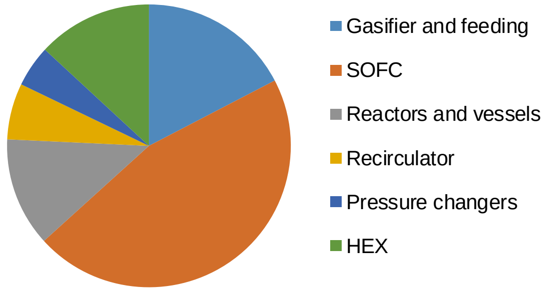

Figure 11. Breakdown of the total purchase cost of the BLAZE plant.

The NPV of the BLAZE plant is at base case conditions -2.8 MEUR. Based on the results obtained above, the sensitivity analysis considers the prices of electricity and biomass, the ISBL and the fixed operating costs, to reach a NPV of at least zero. The effect of the ISBL and the fixed operating costs are more remarkable in NPV terms. NPV equal to zero can be reached if the ISBL goes down from 1.601 MEUR to 0.104 MEUR (i.e. CAPEX of 7,238 EUR/kW – still almost double the target) (with an electricity price of 280 EUR/MWh). In order to reach an electricity price of 100 EUR/MWh (target of BLAZE project) (starting NPV of -3 MEUR, and considering a biomass price of zero), the ISBL should go down to 0.023 MEUR (i.e. 1,600 EUR/kW).

These results point out the importance of the investment and operating costs in BLAZE plant. Particularly important is R&D, to decrease the investment and fixed and variable operating costs. A combination of favourable conditions will be needed for the BLAZE plant to be economically competitive; for instance, a low biomass price will favour a lower electricity price; however, the 100 EUR/MWh target price seems too low.

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7 CONCLUSIONS

The current deliverable summarises the different scenarios evaluations and the optimisation results performed to propose a BLAZE plant layout. From the three recirculation options proposed in the project, the AOG recirculation towards the gasifier combustor was selected as preferred layout for implementation. The optimisation of this layout pinpointed the most important plant variables for plant operation. The electrical and thermal efficiencies can go up to 49 % and 47 %, respectively. However, in terms of costs, the BLAZE plant is still far from its objective.

The final pilot plant implementation decision will come from the adjustment and consideration as baseline of the optimisation results to the strategic decision of the consortium in terms of steam and LPG consumptions, and of course, of practical implementation considerations.

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