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Authors	Norbert N. Vasen (EUBIA), Giulio Poggiaroni (EUBIA), Enrico Bocci
	(USGM), Andrea di Carlo (UNIVAQ), Lex Heijnen (HYGEAR), Jan
	Pieter Outweltjes (SP),
Contributors	
Reviewer(s)	USGM





TABLE OF CONTENTS

1 INTRODUCTION -THE BLAZE CONCEPT	4
2 BIOMASS CHARACTERIZATION	5
3 THE GASIFIER	6
FEEDSTOCK AND BED MATERIAL Pilot plant	6 8
Biomass feeding system	10
Air and steam supply system	10
Gas cleaning and venting systems	11
Gas analysis and tar sampling	11
Start-up phase and test procedure	11
Results and discussion	11
THE GAS CLEANING UNIT	20
DESIGN OF THE CONTAINER	21
GAS UPGRADING UNIT	23
INTEGRATION OF THE TURBOFAN	24
INTEGRATION OF THE LSM	25
5 A NOVEL THERMALLY DRIVEN HIGH-SPEED GAS BEARING SUPPORTED	
SYNGAS COMPRESSOR	26
6 A NOVEL SOFC	31
7 POTENTIAL USE AND PERSPECTIVES	37





1 INTRODUCTION -THE BLAZE CONCEPT

The BLAZE project, funded under Horizon 2020, started in 2019 and has concluded in May 2023 aims at the development of a compact indirectly heated dual bubbling fluidised-bed gasifier (IBFBG: composed of a gasifier within a combustor) integrated with primary sorbents and ceramic candle filters filled with Ni catalysts, high temperature fixed bed sorbents reactors and solid oxide fuel cell (SOFC) including first-of-a-kind heat-driven syngas blower (Fig.1).



Fig. 1.1. BLAZE process scheme

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The technology is developed for a novel CHP with a capacity range from 25-100 kWe (small scale) to 0.1-5 MWe (medium scale) and is characterised by the widest fuel spectrum applicable (forest, agricultural and industrial waste also with high moisture contents, organic fractions of municipal waste, digestate), high net electric (50%) and overall (90%) efficiencies as well as almost zero net GHG and PM emissions.

The project has come to an end and a demonstration pilot plant has been realized in Italy. Promising results and evidence have emerged during the project's activities and this publication is wrapping up the main outcomes. The next section will describe 1) the Biomass characterization activities, 2) the realization of the Gasifier, 3) the development of the gas cleaning unit and the integration components, 4) the development of the novel syngas compressor, 5) The solid-oxide fuel cell and its integration with the system.

2 BIOMASS CHARACTERIZATION

The potential of waste biomass in EU was assessed on the basis of availability, physical properties (low water content and high bulk density), chemical properties (high Calorific Value, high content of volatile substances in order to produce more gas, low ash content, high Carbon to Nitrogen ratio, low Chlorine and Sulphur content), and economic aspects (e.g. cost). Moreover, full characterization of the biomass selected as the representative feedstocks (proximate and ultimate analysis, elements determinations, ignition and burn-out temperatures, ashes characterization) was conducted (see figure below and www.blazeproject.eu/resources). From the analysis of the collected data set all woody and herbaceous biomass feedstocks were usable for gasification with a BFB reactor, since no significant risk of reactor bed defluidisation is expected. However, for most feedstocks, the presence of contents of S and Cl could lead to gaseous products containing S and Cl (e.g. H2S, HCl and alkali halides), the levels of which are too high for immediate use in a SOFC. A first gas cleaning to reduce their concentrations at levels consistent with the SOFC specification needs to be considered for all these biomass feedstocks [1].

The assessment further revealed that corn cobs, black liquor (BL), MSW and digestate are less attractive for gasification in a BFB reactor. Corn cobs and BL were unsuitable due to their rather low ash melting temperatures compared with the typical values adopted in BFB gasification (i.e. slightly above 600 °C vs 800-850 °C) thus leading to a possible reactor block. MSW and digestate appeared as utilizable feedstocks, although at reduced performance due to the significantly lower heating values compared with all the other considered matrices, and due to the higher ash content.

Moreover, their characterization also revealed a rather high content of K, Na, Pb and Zn, which in combination with the high content of Cl could lead to the formation of their respective chlorides, present in the form of vapors in the product gas. KCl and NaCl are known to have a negative effect on SOFC performances; no reference literature is available about PbCl2 and ZnCl2 and their effect on SOFC, however their presence in the producer gas must be considered because of their environmental issues. RDF, MSW and digestate can, on the basis of the ultimate analysis, lead to a producer gas with relatively high contents of H2S and HCl and therefore their formation should be taken into account and properly addressed, since both species are known to have deleterious effects on the stable and long-term functioning of the SOFCs [1].





Feedstock	CATEGORY	Humidity (%- wt, as received)	LHV MJ/kg	Ash %wt, dry basis	S %wt, dry basis	Cl %wt, dry basis	Ash melting T (DT) (°C)
Subcoal	Municipal waste	3,20	21,68	15,60	0,10	1,00	1250,00
Olive pomace pitted	Secondary residues of industry utilising agricultural products	36,30	19,79	5,95	0,06	0,08	1290,00
Sawmill waste	Primary residues from forest	11,20	18,89	0,41	<0.01	<0.01	1300,00
Multi-essence wood chips	Waste from wood	24,50	17,88	1,45	0,02	<0,01	1370,00
Olive Prunings	Secondary residues from wood industries	14,90	17,76	1,55	<0.01	<0.01	1380,00
Almond shells	Secondary residues of industry utilising agricultural products	10,00	17,68	1,31	<0.01	<0.01	1000,00
Swarf and sawdust	Secondary residues from wood industries	6,60	17,14	0,43	<0.01	<0.01	>1385
Wood chips	Primary residues from forest	8,90	16,74	0,54	<0.01	<0.01	>1385
Corn cobs	Agricultural residues	9,00	16,62	3,04	0,03	0,44	645,00
Arundo Donax	Agricultural residues	10,10	16,25	3,43	0,11	0,29	1185,00
1- Wheat Straw (pellets 10 mm)	Agricultural residues	7,60	15,98	9,22	0,05	0,12	1065,00
2- Wheat Straw (pellets 6 mm)	Agricultural residues	7,60	15,40	13,29	0,08	0,21	1135,00
Rice husks	Secondary residues of industry utilising agricultural products	5,20	15,19	14,70	0,02	0,03	990,00
Digestate	Digestate from biogas production	71,20	12,69	25,81	0,97	0,10	1245,00
Black Liquor	Secondary residues from wood industries	20,60	11,20	48,28	0,74	0,12	680,00
Municipal solid waste	Municipal waste	23,00	10,22	47,01	0,20	0,40	1220,00

Table 2.1. Results of the biomass charachterisation acitvities

3 THE GASIFIER

A major activity of the project was the development of an innovative pilot-scale dual bubbling fluidized bed (DBFB) gasifier. In the last two years, the performance of the pilot-scale gasification unit has been studied by varying the operating conditions to select the optimum parameters. Several tests have been carried out to evaluate the effect of auxiliary fuel flow rate, equivalent ratio, steam to biomass ratio (S/B) and temperature on the operating performance of the gasifier. The producer gas was also characterized in terms of dry gas composition, and organic and inorganic contaminants. Furthermore, the efficiency of the hot gas cleaning and conditioning system was evaluated with preliminary tests. Several hundred hours of experiments have been collected.

Feedstock and bed material

Hazelnut shells were used for the gasification tests. The biomass was treated as received (particle size was about 0.5-1.5 cm, see Figure 3.1.







Figure 3.1. Sample of hazelnut shells used for the gasification tests

The results of the proximate and ultimate analysis of the biomass are reported in Table 2.1.

	Ash (wt% _{db})	1.20
Proximate analysis	Volatile matter (wt% _{db})	75.50
	Fixed carbon ^a (wt% _{db})	23.3
	C (wt% _{db})	50.96
	H (wt% _{db})	5.72
Elemental analysis	N (wt% _{db})	0.42
	S (wt% _{db})	0.03
	O ^b (wt% _{db})	41.67
	HHV (MJ/kg) _{db}	19.93
	LHV (MJ/kg) _{db}	18.12

Table 2.1. Characterization of hazelnut shells

^a obtained by difference

^b % O =100 - (% C + % H + % N + % Cl + % S + % ash), dry basis.





Olivine sand supplied by Nuova Cives srl (https://www.nuovacives.com/) was used as bed material. The olivine particle density is equal to 3300 kg/m³, while the bulk density is between 1800-1900 kg/m³. Figure 3.2 reports the particle size distribution.



Figure 3.2. Particle size distribution of olivine

The Sauter mean diameter, calculated by equation 1, is equal to 557 μm :

$$d_{3,2} = \frac{1}{\left(\sum_{i=1}^{n} \left(\frac{x_i}{d_i}\right)\right)} \tag{1}$$

x_i represents the weight fraction of materials with a diameter equal to d_i.

Pilot plant

The dual bubbling fluidized bed gasifier is designed to process up to 20 kg/h of biomass (100 kWth). The flowsheet and a picture of the pilot plant are shown in Figure 3.3 and Figure 3.4 . A schematic representation of the solid bed material circulation scheme and the main dimensions of the reactor are reported in Figure 3.5. A detailed description of the design concept is reported in a previous work¹.

¹ Di Carlo A., Savuto E, Foscolo PU, Papa AA, Tacconi A, Del Zotto L, et al. Preliminary Results of Biomass Gasification Obtained at Pilot Scale with an Innovative 100 kWth Dual Bubbling Fluidized Bed Gasifier. Energies 2022, Vol 15, Page 4369 2022;15:4369. https://doi.org/10.3390/EN15124369.







Figure 3.3. Flowsheet of the pilot plant



Figure 3.4. Picture of the pilot plant







Figure 3.5. Reactor sketch (dimension in mm) highlighting solid bed material circulation

Biomass feeding system

The biomass is fed to the gasification chamber using a system consisting of a hopper and a two-screws conveyor. The feeding system is maintained in an inert atmosphere (with CO_2) and at slightly positive pressure to avoid the backflush of hot gases and prevent the pyrolysis of the biomass before its injection into the reactor. The biomass flow rate is set by the variable-speed motor of the screw conveyor.

Air and steam supply system

The air flow, pre-heated at 200 °C by means of electric heaters, is fed to both the combustor and the gasifier during the start-up phase. During the gasification test, the air entering the gasifier is replaced by steam. A liquefied petroleum gas (LPG) steam generator is used to produce the steam, which is overheated up to 200 °C by an electric heater. The steam is also fed to the two siphons, which are used for the circulation of the bed between the two chambers. During the tests, the pressure inside the gasifier is kept slightly higher than the combustor in order to avoid air leakage.

Calibrated orifices were used to control the flow rates of air and steam.





Gas cleaning and venting systems

The final configuration of the pilot plant, according to the objectives of the BLAZE project, involves the coupling of the gasifier with a vessel containing the hot-gas conditioning system (6 ceramic filter candles of 1,5 m length filled with commercial catalyst). The tests to evaluate the performance of this coupling are also reported in this work. For this reason, a short ceramic filter candle (OD 0.06 m, ID 0.04 m, 0.5 m high) is installed in the freeboard of the gasifier. To simulate the operating conditions of the hot-gas conditioning system, the same residence time of the gas in the catalyst bed was guaranteed. Approximately 1 % (0.18 Nm³/h) of the producer gas is sucked through the filter element using a vacuum pump for composition and tar content analysis. The remaining gas passes through a cyclone and flows to the flare for combustion before being evacuated by the chimney. Also, the flue gas from the combustor passes through a cyclone to reduce particulate content and it is evacuated by the chimney.

Gas analysis and tar sampling

A slipstream of fuel gas passes through the ceramic filter candle and then flows through a series of impinger bottles filled with 2-propanol for tar sampling according to the standard protocol CEN/TS 15439.

The downstream gas (dry and clean) is analyzed in terms of H_2 , CO, CO₂, and CH₄ volume fractions using an online ABB analyzer (Caldos and Uras). In addition, gas bags are collected for offline analysis with an Agilent Micro-GC 990 to quantify the content of H_2S , light hydrocarbons, and other permanent gases (i.e., N_2). The tar samples are analyzed offline with an Agilent 7890 GC-MS to identify tar compounds and calculate their content in the syngas.

Start-up phase and test procedure

During the start-up phase, the reactor is fed with air preheated to 200 °C in both chambers and in the two siphons, as mentioned above. In addition, a 20 kW LPG start-up burner is used in the gasifier to help heating-up the fluidized bed. As the temperature increases, the air flow is reduced, leaving the reactor to operate in a steady fluidization regime. When the temperature in the gasifier increases to approximately 300 °C, the biomass is also fed to accelerate the heating phase through its combustion heat. As the reactor reaches the operating temperature (after about 4 hours), the LPG burner is turned off, and the air fed to the gasifier and siphons is switched to steam. Also, the air flow rate and the auxiliary fuel (LPG) are set to the values chosen according to the operating condition of the combustor.

Reported data refer to the stationary conditions of the plant, that is, after the reactor has reached the operating temperature (750-850 °C) and the flow rates have been set to the values chosen for the specific gasification test. All experimental tests are operated in a day, so the continuous gasification phase (after the start-up phase mentioned above) lasts about 400 min.

Temperature, pressure, feed rates, and gas composition are monitored and recorded continuously during tests by a computer-based data acquisition system. In all tests, the pressure is close to atmospheric conditions.

Results and discussion

Several tests were carried out to evaluate the performance of the gasification unit. Table 3.2 and Table 3.3 show the main operating parameters of the tests conducted at biomass feeding rates of 15 and 20





kg/h, respectively. The steam flow rate is about 13 kg/h in all tests. Therefore, the S/B differs according to the biomass feeding rates (mean value of \approx 1 and \approx 0.8 at 15 kg/h and 20 kg/h, respectively).

The S/B ratio has been calculated according to the following expression:

• Steam to Biomass ratio (S/B):

$$S/B = \frac{\dot{m}_{steam} + \dot{m}_{moist}}{\dot{m}_{Bio,dry}}$$
(2)

Where \dot{m}_{steam} and \dot{m}_{moist} are the mass flowrates of the process steam and of the moisture, respectively, while $\dot{m}_{Bio,drv}$ is the mass flow rate of dry biomass.

As reported in Table 3.2 and Table 3.3**Table 3.3**, the temperature levels in the gasifier and in the combustor chambers always differ by less than 40°C, respectively, demonstrating the efficiency of the heat exchange and circulation of sand between the two fluidized beds (gasifier and combustor).

The bed material circulation for the gasification system used in this work has been already studied by the same authors with a cold model of the innovative DBFB gasifier [36]. The results obtained by means of Lagrangian Particle Tracking showed that the obtainable bed material circulation is higher by a factor of 2–3 than that assuring effective heat exchange between the two chambers, as required by an allothermal gasification process. These results and the heat exchange through the cylindrical contact surface between the two chambers could well explain the small difference between combustor and gasifier temperature, somewhat less than that mentioned by Fercher et al. ²for a dual fluidized bed gasifier composed of two separate reactors (combustor and gasifier).

						8
Test	#1	#2	#3	#4	#5	#6
T _{Gasifier} (°C)	787	800	814	841	846	846
T _{Combustor} (°C)	788	808	815	857	872	849
T _{Freeboard} (°C)	748	734	754	747	740	750
S/B	1.04	0.94	0.97	1.05	0.96	0.91
LPG Flow rate (kg/h)	1.9	1.9	1.7	1.7	1.1	1.1

² Fercher E, Hofbauer H, Fleck T, Rauch R, Veronik G. Two years experience with the FICFB-Gasification process. 10th Eur. Conf. Technol. Exhib., Wurzburg, Germany: 1998, p. 3–6.

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Test	#7	#8	#9	#10	#11	#12	#13	#14	#15	#16	#17	#18
T _{Gasifier} (°C)	782	790	792	799	804	809	815	820	821	822	825	830
T _{Combustor} (°C)	811	801	805	811	825	815	820	827	832	834	859	850
T _{Freeboard} (°C)	732	731	730	724	729	728	744	754	744	730	729	758
S/B	0.80	0.79	0.78	0.77	0.78	0.81	0.96	0.77	0.73	0.79	0.73	0.82
LPG Flow rate (kg/h)	0.9	1.9	1.9	1.9	0.9	1.9	1.9	2.0	1.7	1.9	1.1	1.6

Table 3.3. Main operating conditions of the tests at 20 kg/h biomass feeding rate

The results were evaluated according to the following operational parameters:

• Cold gas efficiency (CGE):

$$CGE (\%) = \frac{Q_{Syngas} \cdot LHV_{Syngas}}{\dot{m}_{Bio} \cdot LHV_{Bio} + \dot{m}_{LPG} \cdot LHV_{LPG}} \cdot 100$$
(3)

• Carbon conversion (C_{conv}):

$$C_{Conv} (\%) = \frac{\sum_{i=CO,CO_2,CH_4,tar} mol_i}{mol_{C,Bio}} \cdot 100$$
(4)

Where mol_i are the moles of C contained in the CO, CO₂, CH₄ and tar that exit with the producer gas while $mol_{C,Bio}$ are the moles of C that enter with the biomass.

• Water conversion (H₂O_{Conv}):

$$H_2 O_{Conv} (\%) = \frac{\sum_{i=H_2, CH_4, tar \, mol_i - mol_{H2, bio}}{mol_{steam} + mol_{moist}} \cdot 100$$
(5)

Where mol_i , in this case, are the moles of H₂ contained in H₂, CH₄ and tar that exit with the producer gas, $mol_{H2,bio}$ are the moles of H₂ that enter with the biomass (see elemental analysis in Table 1) and finally





 mol_{steam} and mol_{moist} are the moles of H₂O that enter with the process steam and the moisture, respectively.

The amount of gas produced was calculated using the expression from Fercher et al³.

Gas yield
$$\left(Nm^3/kg_{Bio,dry-ashfree}\right) = \frac{33.333}{(71.375 - 0.05207 \cdot T)}$$
 (6)

where T is the temperature (°C). Equation 6 allows to calculate the amount of syngas produced, which is close to the results obtained and reported in a previous work⁴.



Figure 3.6.a

³ Fercher E, Hofbauer H, Fleck T, Rauch R, Veronik G. Two years experience with the FICFB-Gasification process. 10th Eur. Conf. Technol. Exhib., Wurzburg, Germany: 1998, p. 3–6.

⁴ Di Carlo A., Savuto E, Foscolo PU, Papa AA, Tacconi A, Del Zotto L, et al. Preliminary Results of Biomass Gasification Obtained at Pilot Scale with an Innovative 100 kWth Dual Bubbling Fluidized Bed Gasifier. Energies 2022, Vol 15, Page 4369 2022;15:4369. https://doi.org/10.3390/EN15124369







Figure 3.7.b Temperature values measured during the start-up and gasification phases (a) and syngas composition measured

during a gasification test (b) at 20 kg/h biomass feeding rate

Figure 3.7.a shows the temperatures measured in the two beds and in the freeboard during a typical startup phase and gasification test. Furthermore (see Figure 3.7.b), the gas composition during the sampling phase is reported (reactor at the steady state). As shown in Figure 3.7.a, temperatures are constant during the gasification tests. The gasification temperature was maintained constant at a desired value varying the air flowrate (between 34 and 40 kg/h) and the auxiliary fuel flow rate in the combustor (between 8 and 18 l/min in gas phase). The influence of air flow on the combustor was studied to verify both the leakages between the two chambers and the reactor's temperature.

Figure 3.8 shows the compositions of the syngas as a function of the operating temperature for the two biomass feeding rates. Hydrogen content increases and CO_2 decreases with the temperature at 20 kg/h of biomass feeding rate, probably due to the higher reactivity of the char at a higher temperature. Furthermore, it is slightly higher at 15 kg/h compared to 20 kg/h due to the higher steam-to-biomass ratio.







Figure 3.8. Gas composition at different gasification temperature values: a) 15 kg/h and b) 20 kg/h biomass feeding rate

As shown in Figure 3.8, the tests conducted with similar conditions give comparable results, demonstrating the repeatability of the tests, and, therefore, the reliability of the system.



Figure 3.9. Tar concentration in the produced syngas





In the range of temperature values investigated, the tar concentration in the gas is not significantly affected by the operating temperature. For this reason, Figure 3.9 reports the average concentration values of tar compounds detected varying the operating temperature for the two biomass feeding rates (GC-MS analyzer was used to identify and quantify the tar compounds). The tar content obtained was 12.4 g/Nm³ and 15 g/Nm³ for biomass feeding rates of 15 kg/h and 20 kg/h, respectively. The lower concentration observed at 15 kg/h can be ascribed to the higher S/B ratio. The tar content of syngas is too high for its direct use in downstream processing/operation; however, tars can be significantly reduced with gas conditioning systems.

The Micro-GC analyses on syngas sampled with the gas bags indicate a nitrogen content of approximately 5% in all tests, confirming that the gas leakage from the combustor to the gasifier is minimum and the separation of the gas streams between the two chambers is effective. Furthermore, the H₂S content was analyzed. The results showed a content between 50 to 100 ppm.

The particulate matter was recovered from the two cyclones (see Figure 3.3). The content is about 15 g/Nm³ and 7 g/Nm³ in the flue and fuel gas, respectively. Based on the results obtained in a lab-scale gasification plant, the particulate content of the producer gas will be completely abated by using the ceramic filter candle.



Figure 3.10. LHV and Cold gas efficiency as a function of temperature: a) 15 kg/h; b) 20 kg/h biomass feeding rate

As shown in Figure 3.10, the syngas produced for all tested conditions has a lower heating value (LHV) of approximately 10.6 MJ/Nm³, more than double of the value obtained with air gasifiers (4-5 MJ/Nm³).

The CGE shows an increasing trend with the temperature at a biomass feeding rate of 15 kg/h. On the other hand, tests at 20 kg/h show a variable trend of CGE with temperature. These results are related to unfavorable operating conditions (lower excess of air and reduced heat exchange between chambers). As reported in Table 2 and 3, the increase in LPG flow rate in some tests does not lead to a rise in the





gasification temperature. The best results obtained at higher temperatures are related to the optimal LPG/biomass ratio and with an air excess of 30%. The optimal LPG/Biomass ratio in terms of energy input is approximately 20%, which aligns with the results reported with different dual fluidized bed gasifiers⁵.

The results, reported in Figure 3.11, show that carbon conversion is not significantly affected by the temperature at a biomass feeding rate of 20 kg/h, while a growing trend is obtained at 15 kg/h. The carbon conversion results were very similar in all tests with a mean value of (77.2 ± 2.8) %. The value, calculated according to equation (4), doesn't take into account the light hydrocarbons (C2-C6) not measured in this work. Larsson et al.⁶, in previous work, evaluated the typical gas composition that can be obtained during biomass steam gasification at industrial scale plants and showed that these compounds represent approximately 2-4 % of the total gas produced.

The water conversion results were higher in the tests at 20 kg/h of biomass feeding rate than at 15 kg/h (mean value of 15.4 % and 13.7 %, respectively). As for the parameter analyzed above, the results are strictly related to the operating conditions that allow to obtain the highest gasification temperature. In particular, it is evident that at 20 kg/h increasing the temperature, the water-gas reaction is enhanced, then the water conversion increases producing more CO and H₂, as reported in Figure 3.8. Furthermore, the water conversion is strictly related to the S/B ratio. The values obtained are comparable to that reported by Hofbauer and Rauch⁷.



Figure 3.11. Carbon and water conversion as a function of temperature: a) 15 kg/h; b) 20 kg/h biomass feeding rate

⁵ Wilk V, Schmid JC, Hofbauer H. Influence of fuel feeding positions on gasification in dual fluidized bed gasifiers. Biomass and Bioenergy 2013;54:46–58. https://doi.org/10.1016/J.BIOMBIOE.2013.03.018.

⁶ Larsson A, Kuba M, Berdugo Vilches T, Seemann M, Hofbauer H, Thunman H. Steam gasification of biomass – Typical gas quality and operational strategies derived from industrial-scale plants. Fuel Process Technol 2021;212:106609. https://doi.org/10.1016/J.FUPROC.2020.106609

⁷ Hofbauer H, Rauch R. Stoichiometric Water Consumption of Steam Gasification by the FICFB-Gasification Process. Prog Thermochem Biomass Convers 2008:199–208. https://doi.org/10.1002/9780470694954.CH14.





As mentioned, the raw syngas shows a high tar content, in particular, exceeding the requirements for coupling the gasifier with the SOFC⁸. For this reason, preliminary tests with a filter candle filled with commercial Ni-based catalyst installed in the freeboard of the gasifier were carried out to evaluate the in situ hot gas cleaning, conditioning performance and the behavior of the catalyst during extended tests. The commercial steam reforming catalyst was supplied by Johnson Matthey. The pellet catalyst with a diameter and height of 3 mm contains between 10 wt.%. to 15 wt.%. of Nickel. This catalyst was previously tested inside a short candle inserted in the freeboard of a bench scale fluidized gasifier and showed a good performance regarding tar conversion: for this reason, that pellet catalyst was also selected for this work. The gas hourly space velocity (GHSV) for these tests was maintained constant at around 1000 h⁻¹.

The results in Figure 3.12 show that the tar concentration tends to decrease with the test duration, reaching a value of about 2 g/Nm^3 . This can be ascribed to the fact that the catalyst is not initially activated but a certain period is required for its reduction by means of the same syngas. The reduction of tar content confirms the activity of the commercial catalyst used for the steam reforming reactions. The comparison of the syngas composition obtained in the long run test (with the catalyst in the ceramic candle-Figure 12) shows a higher content of H₂ and CO₂ and lower content of CO (see Figure 3.7 for comparison), highlighting the effect of the water gas shift (WGS) reaction.

Finally, the gas yield obtained in the tests with catalyst has been calculated by mass balances, starting from the results obtained without catalyst and taking into account the reduction of tar and the WGS reaction. The estimated mean value for the long test run is $\approx 1.4 \text{ Nm}^3/\text{kg}_{\text{Bio,dry,ash free}}$.



Figure 3.12. Time evolution of tar concentration in the syngas using a filter candle filled with catalyst

⁸ Ouweltjes JP. "D3.2 BLAZE Project Report summarising the literature review" for Selection of Representative Syngas Compositions including Organic and Inorganic Contaminants. 2019.



Figure 3.13. Time evolution of syngas composition using a filter candle filled with catalyst (data of test 8 not available)

As reported in Table 3.2 and Table 3.3.3, the temperature of the freeboard is always lower than that of the gasifier (up to 100 °C lower); for this reason, further tests will be carried out to increase the temperature in the freeboard and improve the performance of the catalyst. Tests conducted with a laboratory-scale gasifier and companion numerical simulations already demonstrated the reduction of tar content, which remains within limits imposed for the SOFC by using this catalyst at 850 °C⁹.

THE GAS CLEANING UNIT

The Gas Cleaning Unit (GCU) consists of a series of sorbent vessels and a tar reformer to reduce the concentration of the Chlorine and Sulphur compounds and the tars to levels the SOFC can tolerate. The chlorine and sulphur compounds are taken out of the gas stream by using dedicated sorbent at a temperature level of about 400°C. Tar cracking occurs at a slightly higher temperature level using special tar catalyst material. To accommodate the various temperature levels heat exchangers are part of the Gas Cleaning Unit.

The GCU prototype is built inside a 40 ft sea container and consists of the following sections:

⁹ Papa AA, Savuto E, Carlo A Di, Tacconi A, Rapagnà S. Synergic Effects of Bed Materials and Catalytic Filter Candle for the Conversion of Tar during Biomass Steam Gasification. Energies 2023, Vol 16, Page 595 2023;16:595. https://doi.org/10.3390/EN16020595.

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1. A Gas Upgrading section, containing all high temperature components required for the proper operation of the SOFC module, such as: sorbents for gas impurities, tar reformer, heat exchangers etc.;

- 2. A large stack module (LSM) with a nominal power output of 25 kWe, and;
- 3. A steam driven turbofan.

Design of the container

The following pictures show the exterior and the interior of the CHP housing. The exterior of the container is shown in Figure 4.1 and Figure 4.2; a view of the containers' interior is given in Figure 4.3 to Figure 4.11.



Figure 4.14: The outside of the housing showing the vent pipes and the "side 1 interface" panel



Figure 4.15: Opposite exterior of the housing showing "side 2 interface" panel





Utility connections provide the connections to flush the system and to supply compressed gas to the pneumatic valves, Figure 4.16.



Figure 4.16: Utilities connected to "Side 2 interface" panel. These provide gas to flush the system and air to operate the pneumatic valves

The control cabinet is designed and build by Hygear, see Figure 4.17. Cable trays are placed as well as the brackets to support the trays, instrumentation and piping.



Figure 4.17: part of electrical cabinet





Gas upgrading unit

The gas from the gasifier will be upgraded from impurities before entering the Turbofan. Three high-temperature resistant valves will separate the GCU from the Gasifier, Figure 4.18.



Figure 4.18: High temperature interface valves.

The gas is cleaned with two high temperature sorbent vessels and one TAR reformer. These contain thermocouples and band heaters to maintain the correct temperature, Figure 4.19.



Figure 4.19: High temperature sorbent vessels (left) and tar reformer (right) to remove gas impurities. Band heaters provide equal heat transfer to the vessel





Integration of the Turbofan

To suck the gas from the gasifier a steam driven compressor is used. A 3D printed model of the Turbofan is mounted in order to fit the piping and bracket, Figure 4.20. To drive the compressor, the heaters will increase the temperature of the incoming steam. Coriolis mass flow meters will provide accurate measurement regarding steam mass flow.



Figure 4.20: 3D printed module of the turbofan. The model is used to fit the actual design.



Figure 4.21: Closeup of steam and fuel piping

The 3D model will be changed with the actual prototype after completion of the lab test by EPFL and shipping of the container to WT.





Integration of the LSM

The LSM is placed in the back side of the container and connected to the LSM, Figure 4.22. Heaters provide the correct temperature for the fuel inlet and air inlet, Figure 4.23.



Figure 4.22: Close up of the LSM unit with the hot ventilation air exhaust



Figure 4.23: Close up of the LSM hot air heaters (left) and the fuel heater (right)

After separating the hydrogen from the syngas the gas is cooled down and water condense is collected inside a water knock-out, Figure 4.24. The gas is then recirculated and mixed with the incoming syngas from the gasifier.







Figure 4.24: Anode off gas cooler. Before recirculation, the gas is cooled down and water is taken out.

5 A NOVEL THERMALLY DRIVEN HIGH-SPEED GAS BEARING SUPPORTED SYNGAS COMPRESSOR

Over the course of the project, the concept of the hot gas anode off-gas recirculation has been replaced with the design and development of a novel thermally-driven syngas compressor that couples the biomass gasifier with the SOFC stack. The reason is the discovery of a negative impact of AOG recirculation in a syngas-fueled SOFC system on the electrical efficiency due to fuel dilution¹⁰.

Further, the lack of off-the-shelf small-scale pressurized screw feeders for the gasifier, as well as the difficulties in the pressurization of the latter (leakages, fluctuations in pressure), forced a change of the demonstrator's layout. Hence, for the coupling of the biomass gasifier with the SOFC, both operating at ambient pressures, a syngas compressor to overcome the pressure losses in the pilot plant is required. To lower the energy consumption for the syngas compression and thus increase the pilot plants overall system efficiency, a novel thermally driven high-speed high temperature syngas compressor supported on gas bearings has been developed at EPFL LAMD.

Experimentally validated 0D to 3D design procedures coupled with SOFC-gasifier system level process modeling as well as gas bearing and rotor optimization using an artificial neural network (ANN) are employed to achieve the final design of the compressor-turbine unit (CTU). The radial compressor impeller consists of 18 backward-curved main and splitter blades, having been designed for an inlet syngas mass flow of 18.23 kg/h at 350°C and 0.81bara. The full-admission cantilever steam turbine, inspired by the

¹⁰ Pérez-Fortes, M., He, V., Nakajo, A., Schiffmann, J., Maréchal, F. and Van herle, J., 2021. "Techno-Economic Optimization of an Integrated Biomass Waste Gasifier-Solid Oxide Fuel Cell Plant". Front. Energy Res. Vol.9 2021





design of Wagner et al.¹¹, uses steam at design inlet pressures of 3.5bara and temperatures at 525°C to drive the syngas compression at the compressor side. The turbine design pressure ratio is 2 (thus design outlet pressure at 1.75bara). However, an expansion to ambient pressure is possible to achieve higher turbine powers (to overcome bearing losses and provide compression power) at the cost of efficiency (isentropic total-to-static efficiency of ~33% and turbine power of 2768W for 1bara turbine outlet pressure from CFD simulations).

The CTU is supported on ANN-optimized dynamic gas bearings, namely two herringbone-grooved journal bearings and a two-sided spiral-grooved thrust bearing, that allow an oil-free, low-wear, high-temperature and high-speed operation (nominal rotational speed of 210krpm) of the CTU. A validated 1D heat model based on Olmedo et al.¹² has been established for the CTU system, where the results indicate a build-up of heat within the CTU system due to the windage losses at the gas bearings, thus leading to unwanted thermal gradients, increasing the risk of material and operational. To mitigate this, the housing and thus the rotor are actively flushed with steam at 1.25bara and 412°C. The nominal operating conditions as determined from CFD simulations and the used materials are shown in

Syngas compressor parameters	Design point
Syngas flow rate, kg/h	18.23
Compressor inlet temperature, C	350
Compressor outlet temperature, C	426
Compressor inlet static pressure, bara	0.81
Compressor outlet static pressure, bara	1.16
Delta P compressor, mbar	350
Compression power, W	730
Isentropic total-to-total full stage compression	0.75
efficiency	
Compressor material	Ti Grade 5 with high-temperature
	anti-corrosive coating
Turbine inlet temperature, C	525
Turbine outlet temperature, C	429
Turbine inlet static pressure, bara	3.5
Turbine outlet static pressure, bara	1.75
Turbine steam mass flow rate demand, kg/h	72
Turbine power, W	2324
Isentropic total-to-static full stage expansion efficiency	0.49
Turbine material	Ti Grade 5 with high-temperature
	anti-corrosive coating
Gas bearings steam inlet temperature. C	412
Gas bearings steam inlet pressure, bara	1.25
Gas bearings steam inlet mass flow, kg/h	8

 ¹¹ Wagner, P.H., Van Herle, J. and Schiffmann, J., 2020. "Theoretical and experimental investigation of a 34 Watt radial-inflow steam turbine with partial admission". ASME Turbo Expo 2020
 ¹² Olmedo, J. E. Liu, W. Gijka, K. Schiffmann, J. 2023. "Thermal management for gas lubricated high-speed."

¹² Olmedo, L.E., Liu, W., Gjika, K., Schiffmann, J., 2023. "Thermal management for gas lubricated, high-speed turbomachinery". Applied Thermal Engineering Vol.218 2023

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Rotational speed, rpm | Windage losses, W Rotor and bushing material |

210 000 960 Tungsten carbide

Table 5.4.





Syngas compressor parameters	Design point
Syngas flow rate, kg/h	18.23
Compressor inlet temperature, C	350
Compressor outlet temperature, C	426
Compressor inlet static pressure, bara	0.81
Compressor outlet static pressure, bara	1.16
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Turbine inlet temperature, C	525
Turbine outlet temperature, C	429
Turbine inlet static pressure, bara	3.5
Turbine outlet static pressure, bara	1.75
Turbine steam mass flow rate demand, kg/h	72
Turbine power, W	2324
Isentropic total-to-static full stage expansion	0.49
efficiency	
Turbine material	Ti Grade 5 with high-temperature
	anti-corrosive coating
Gas bearings steam inlet temperature, C	412
Gas bearings steam inlet pressure, bara	1.25
Gas bearings steam inlet mass flow, kg/h	8
Rotational speed, rpm	210 000
Windage losses, W	960
Rotor and bushing material	Tungsten carbide

Table 5.4: Nominal operating conditions of the steam driven syngas compressorSvngas compressor parametersDesign point

Design choices have been made such as to achieve a high feasibility, manufacturability and robustness, as well as a wide operating range. Examples are (1) the design of an unshrouded compressor impeller with constant blade thicknesses to allow easier manufacturing via milling processes; (2) the design of oversized, tunnel-shaped volutes as well as setting the design point of the compressor for the mass flow at the lower end of the range (range: 18.23 - 24.3 kg/h) to widen the operating range and increase robustness against fluctuations of operated conditions; or (3) the consideration of manufacturing deviations of $\pm 2\mu$ m for the optimization of the gas bearings to increase its robustness towards manufacturing imprecision. The final CTU system and main specifications are summarized in Figure 5.1 and Table 5.5.







Figure 5.25: Novel CTU system design

The CTU has been investigated theoretically and numerically using CFD and FEA tools, where results show a good agreement between theoretical models and CFD simulations (difference in isentropic efficiencies by ~3%-points at design points).

Compressor impeller	Parameters	Turbine impeller	Rotor	Rotor
Inlet shroud radius	7.5		Inlet	outlet
r _{in,shroud} , mm		Radius r, mm	10.5	8.5
Inlet hub radius	2.71	, Dlada baight b	27	2.4
r _{in,hub} , mm		Blade height b,	2.7	3.4
Outlet radius rout	19	mm		
mm	10	α (tangential)	25°	116.9°
Outlet blade height	1.7	β_{blade} (tangential)	39.2°	138.3°
b _{out} , mm	nm		Q	9 1°
N° of blades	9 main & 9	blace turning	9.	5.1
	splitter	N° of blades		47
Blade thickness, mm	0.3 (const.)	Blade thickness,	0.25	0.175
α_{in} (tangential)	90° (swirl-free)	mm		
β_{out} (tangential)	65°			

Table 5.5: Compressor	and turbine	impeller main	geometrical	parameters
Tuble 3.5. compressor	una carbine	impener main	Scomethear	parameters

For the first experimental validation of the models, investigation of the start-up and shut-down behaviour and balancing of the CTU, a test rig has been established at EPFL LAMD, allowing the operation under partial admission using overheated steam or pressurized air at the turbine side, and ambient or hot air at





the compressor side. The CTU is controlled over rotor-speed monitoring and adjustment of the steam flow to the turbine, taking into account the operating and material limits.

The start-up and shut-down behaviour of the unbalanced vertically placed CTU (compressor up, turbine down) using pressurized air at 20°C and 7.8bar, have shown a minimum required turbine volume flow of >200l/min to achieve lift-off, and a turbine volume flow of ~130l/min for controlled and safe touchdown at ~11.1krpm. With the unbalanced CTU, speeds up to 18krpm have been experimentally tested. For safe operation at higher speeds, balancing is required. Further, static and dynamic high temperature tests have to be conducted to experimentally ensure a good alignment of the parts and suitable interference press fits between rotor and the connecting plugs, as well as to validate the established 1D heat model and FEA results.

For the full characterization of the CTU, full-admission tests with syngas and steam as well as the start-up and shut-down procedure using forming gas have yet to be conducted at the pilot plant. First compressor maps for start-up and shut-down conditions using forming gas (80% N2, 20% H2) at 1.079 bar have been generated in discussion with the project partners, indicating that an operation outside of surge is possible under these conditions. The P&ID of the syngas-compressor section in the pilot plant as well as the full start-up, shut-down and testing procedure on-site are to be finalized, using the experimental results obtained in in-house tests.

The developed CTU and the concept of converting waste heat to compression power (and electrical power, if a generator is added), as well as the obtained theoretical, numerical and experimental results are not only applicable to the BLAZE plant, but can also be extended and applied to other small-scale high-temperature energy systems (such as gas turbine engines, organic Rankine cycle turbines, high-temperature heat pumps), combined cycles (such as SOFC-GT hybrid systems, as extensively discussed by He et al.¹³), to the mobility sector (in particular, turbochargers) or district heating systems (in particular, by replacing expansion valves to valorise expansion power to compression power and/or electricity). The novel highly integrated design process can also be used by engineers and scientists in similar domains. In fact, a publication on the detailed design methodology and numerical results is currently in progress.

¹³ He, V., Gaffuri, M., Van herle, J., Schiffmann, J., 2023. "Readiness evaluation of SOFC-MGT hybrid systems with carbon capture for distributed combined heat and power". Energy Conversion and Management Vol.278 2023





6 A NOVEL SOFC

SolydEra, previously known as SOLIDpower, has manufactured a 25 kWe Large Stack Module (LSM) to be used in the BLAZE project. The specifications of the LSM, depicted in Figure 6.1, are as follows:

- Power output 25 kWe, integrating 4 stacks of 6.5 kWe
- Fuel: hydrogen or steam-reformed methane, with a maximum convertible H_2 flow of 280 slm, or a maximum convertible CH_4 flow of 70NI/min, at 80% fuel utilization
- Oxidant: Air, with a maximum tolerated flow 5600 slm
- Operating temperature » 750 °C (inlet 690, exhaust 790 °C)
- Maximum stack electrical efficiency: 50%
- Dimensions: 1699 x 792 x 1385 mm
- Weight: 1505 kg

The unit has been designed to operate with different fuels, including producer gas. Ever since it came available, exploratory tests have been carried out under different operating conditions, including tests under hydrogen, reformate, and also tests where the LSM was used as an electrolyzer, which proved its reliability.



Figure 6.1. The Large Stack Module (LSM)

A big unknown was how the LSM would respond to contaminants that may exist in producer gas. For that, a literature was carried out in an early phase of the project. This was followed by extensive testing of SOFC single cells and short stacks under representative conditions, with the aim to understand the impact of producer gas contaminants and which contaminant levels would still be acceptable.





From the literature review, which was carried out by SolydEra, it was concluded that the longevity of the LSM is likely at risk if no further gas cleaning will take place. Particularly problematic gas compounds are tars (in particular 2-ring tars and higher), sulphur, halogens (particularly chlorine), and alkali metals (particularly sodium and potassium).

ENEA carried out a large series of single cell tests, the cells having an active area of 2 cm² and being provided by SolydEra, with the aim to perform a mechanistic study to the impact of the problematic contaminants, where in some cases tests were repeated to verify reproducibility of the deactivation phenomena. A dedicated test bench was built and equipped with tools to perform characterization by means of electrochemical impedance spectroscopy, see Figure 2. Initally, tests were carried out at 0.4 A/cm² current density under clean gas compositions, including H₂/N₂ mixtures, H₂/H₂O mixtures, CO/CO₂ mixtures, and synthetic, more complex gas mixtures (H₂, CO, CO₂, H₂O, CH₄) to understand the electrochemical response of the SOFC anode, O₂/N₂ synthetic mixtures to understand the response of the soFC anode, O₂/N₂ synthetic mixtures to deeply understand the different electrochemical processes governing the SOFC performance, in order to deeply understand the baseline performance of the cells. By means of DRT (distribution of relaxation times), a break up of the different contributions in the EIS spectra was possible, revealing peaks denoted to mass transport losses, ionic diffusion losses, ohmic losses, and charge transfer losses within the SOFC, see Figure 3.



Figure 6.2. SOFC test bench realized at ENEA.







Figure 6.3. Example of the results obtained by DRT analysis, indicating the different electrochemical losses as a function of the perturbation frequency imposed to the cells during the EIS measurements

In this example, the oxygen partial pressure of the gas mixture supplied to the SOFC cathode was varied. It was further shown during a 150 h test that the SOFC performance under clean (synthetic) producer gas was stable.

Subsequently, the tolerance to the typical contaminants In producer gas was assessed:

- Tests under producer gas including 60 ppm or 180 ppm toluene (the most abundant light tar component in producer gas) showed that this contaminant did not have an immediate negative effect on the SOFC overall performance, though there were some indications that higher toluene levels could affect the performance by slowly clogging up pores in the SOFC anode, thus limiting the fuel distribution in the electrode microstructure.
- Tests under producer gas including 5 or 15 ppm naphthalene (the most abundant heavy tar component in producer gas) revealed that this tar compound did not cause problems at a level of 5 ppm, whilst it had an immediate negative effect on the SOFC performance at 15 ppm. From the EIS analysis, it was speculated that this was due to deactivation of the nickel particles in the SOFC anode with the tar due to its sluggish conversion kinetics at the SOFC operating temperature, perhaps being slightly alleviated by improved electrical condctivity within the electrode due to the deposition of carbon species.
- Tests under producer gas including 1 and 3 ppm H₂S revealed deactivation at both contaminant levels, the deactivation at 1 ppm H₂S being low and gradual, whilst at 3 ppm H₂S a rapid initial deactivation, followed by gradual further degradation was observed. From EIS measurements it followed that it was mainly the SOFC anode charge transfer resistance increase, due to adsorption of sulphur on the nickel catalytically active sites.





- Tests under producer gas including both H₂S and toluene revealed that in such case immediate deactivation would take place, even in presence of 1 ppm H₂S and 60 ppm toluene, with EIS analysis revealing a degradation very similar to what was observed in the presence of 3 ppm H₂S without toluene. Surprisingly, at higher contaminant levels, no detrimental degradation was observed, which suggests a competition between the poisoning effects of both contaminants, where toluene perhaps alleviates the poisoning effect of H₂S.
- Tests under producer gas including both H₂S and naphthalene showed that at low contaminant concentrations, also in this case the poisoning effect of H₂S was perhaps alleviated by the presence of naphthalene, whilst under H₂S alone, the cell in question rapidly degraded. At high contaminant levels, rapid initial degradation was observed, after which the performance stabilized. As from tests with the individual contaminants it followed that both contaminants cause a similar electrochemical response, it was unclear which contaminant was responsible for the initial degradation.
- A test under producer gas including both low concentrations of toluene and naphthalene showed no apparent deactivation during a 120 h test.
- A test under producer gas including toluene, naphthalene and H₂S showed that in that case even at low contaminant levels initial deactivation took place, after which the voltage stabilized. This suggests that the observed mitigation effect of tarts towards H₂S poisoning is perhaps related to a limited range of contaminant levels, different for different tar species, while above this range the tars can worsen the impact of H₂S.

Postmortem analysis on the tested samples revealed no clear signs of degradation nor the presence of contaminants that might exist in producer gas, suggesting that the poisoning takes place through adoption rather than physicochemical alteration of the cell materials.

At EPFL, short stack tests were carried out, the test bench being shown in Figure 6.4. In contrast to the tests at ENEA, where it was the aim to perform mechanistic studies with only a limited test time per cell at EPFL the focus was more on testing the durability under producer gas with a selected amount of contaminant levels. The short stacks, that were provided by SolydEra, comprised 6 repeating elements, each repeating element accommodating one 80 cm² SOFC component. Prior to starting the test campaign at EPFL, it was confirmed by SolydEra that the performance of such short stack is representative for the large stacks housed in the LSM.







Figure 6.4. Picture of the adapted test-bench showing the set-up dedicated for testing the new SOLIDpower stack design. The red arrow indicates where the stream containing the impurities will be fed.

A solution had to be found for the generation of H_2S since pre-calibrated H_2S/H_2 gas mixture could not be used because of laboratory safety restrictions. It was decided to use an organic sulphur compound as a precursor that would be decomposed in presence of hydrogen to form H_2S in-situ.

A first short stack was operated for 9000 h under contaminant-free producer gas, giving a reference baseline at 0.4 A/cm² current density. The average degradation rates were around 0.4% per kh for cells 1 to 5, but 3.3% per kh for cell 6. EIS measurements revealed that this cell likely suffered from contact issues.

A second short stack was tested for 7000 h with producer gas including contaminants. Different levels of sulphur and toluene were injected during the last 2000 h to check the effect of the contaminant on the performance and degradation. The followed approach was to increase gradually the sulphur by 0.2 ppm steps from 0 to 1ppm and for each step superpose toluene by steps of 20 ppm up to 60. The stack was maintained under polarization at 0.4 A/cm² at 750°C and electrochemical impedance spectroscopy (EIS) measurements were performed between each sequence. In addition, the inlet and outlet gas compositions of the stack were measured using a gas chromatograph (GC) equipped with a FPD detector for S-compounds, a FID detector for hydrocarbons and a TCD detector for the permanent gases.

Regarding the impact of H₂S, a similar effect was observed as during the first test, with an effect already at 0.2 ppm H₂S. The observed voltage behavior is typical of sulphur poisoning: an initial voltage drop attributed to surface coverage of the nickel in the SOFC anode, followed by stabilization during which the degradation rate decreases. The relation between the H₂S concentration and the voltage drop follows the trend based on Temkin-like adsorption isotherms. The addition of toluene up to 180 ppm did not show





any significant impact on the performance of the stack. A slight improvement of the conversion impedance in the data was observed with increasing toluene level but this is most likely due to the increase of the hydrogen feed which is the carrier gas for toluene and was not compensated for.

A third short stack test aimed for testing once more the impact of H_2S and toluene at 0.4 A/cm² but up to higher contaminant concentrations. In addition, the impact of HCl will be evaluated, which will be done at the end of the test. At the time of writing, the stack was operated for 5500 h. After obtaining the baseline performance with clean producer gas, the impact of both sulphur and toluene was investigated up to 4 ppm H2S and 400 ppm toluene, see Figure 6.5, indicating that sulphur has the biggest impact on performance, with no stabilization in cell voltage beyond 2 ppm, whilst toluene did not affect the performance. Subsequent operation with clean producer gas led to full recovery of the cell voltages.



Figure 6.5. Impact of H_2S and toluene during the third short stack test at EPFL.

In conclusion, it can be stated that among the different contaminants that were evaluated, H₂S and naphthalene are the most problematic, with both having a similar poisoning effect of the electrochemically active sites in the SOFC anode, whereas toluene would be less problematic up to 400 ppm, at least for the test periods considered. The tests at EPFL indicate that testing cells in a stack arrangement, using representative cell sizes, provides the best information about how deep the producer gas has to be cleaned, whereas the tests of (smaller) single cells, which was the approach at ENEA, offers the quickest and most versatile approach in mechanistic studies to the interaction between the different contaminants on producer gas. The mechanistic studies indicate that a synergistic, alleviating effect between the contaminants could occur, albeit at small and rather unpredictable contaminant concentrations.





7 POTENTIAL USE AND PERSPECTIVES

The market analysis performed in WP8 by EUBIA has examined the potential of a new CHP technology, proposed by the Blaze project and based on gasification and solid oxide fuel cells (SOFC) that can work in the range of 100 – 5000 kWe. This analysis shows that this technology has significant potential to become a key player in the energy production in the EU.

The main advantage of biomass CHP systems based on gasification and SOFC is their high efficiency and the potential to make the EU not only more independent on energy import, but also fulfilling the goals of both Renewable Energy & Energy Efficiency. These systems can achieve total efficiency levels of up to 90% (from which not less than 50% electricity), significantly higher than conventional systems. The flexibility of this technology is another advantage, as it can work with a wide range of biomass types, making it a viable alternative for much more % of the fossil fuel demand.

The binding targets for Renewable Energy and Energy Efficiency in the European Union are driving the demand for CHP systems that are based not on foreign fossil fuels but on biomass gasification and SOFC. The EU targets are ambitious and reaching them seems impossible if we go on as usual. Therefore, the EU needs a solution that tackles several obstacles simultaneously and opens new markets.

This document explains also that the high cost and complexity of this technology may limit its adoption by smaller organizations and individuals. To overcome these challenges, manufacturers and suppliers of CHP systems based on gasification and SOFC need to focus on improving the efficiency of these systems, simplifying its use and reducing their costs.

As a new technology, we have to find early adopters to create successes that can be easily transposed to other users. Therefore, three types of activities have been identified that may benefit from this technology in a significant way and will set an example for many market players of the same sector. This document gives practical steps to identify these three groups and starting lighthouse projects.

The market analysis concludes that the potential benefits of CHP systems based on gasification and SOFC outweigh the challenges. The increasing demand for decentralized energy solutions, the quest for fuel independence and the need for reliable backup power sources are expected to drive the growth of this market. Manufacturers and suppliers of CHP systems based on gasification and SOFC need to invest in research and development to improve the reliability and durability of these systems to build trust and confidence in the technology.

Overall, the market analysis suggests that the new CHP technology based on biomass gasification and SOFC that can work in the small/medium power range has the potential to become a game-changer in the energy supply, but it will require the support and investment of both the private and public sectors to realize its full potential.