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Techno-Economic Analysis of Hydrogen and Oxygen Production Plant from Medical Waste for Hospitals in Jordan

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ABSTRACT

Jordan faces significant challenges in terms of energy generation due to a heavy reliance on energy import from neighbouring countries. The country also grapples with waste management issues stemming from various sources, particularly medical waste, despite the economic crisis that the country is going through. Furthermore, an accident occurred in government hospital due to oxygen shortages. To address these concerns, this investigation explores the possibility of self-sufficient production of oxygen to cover hospitals needs in Amman, Jordan, additional to the production of hydrogen to achieve self-sustained hospital operation in terms of power and heat requirement through the utilization of hospital medical waste (MW). Data of MW in Amman city were obtained from the Ministry of Health analysis, including the available supply, proximate and ultimate analysis. A new model using Aspen Plus software was developed to investigate the techno-economic and environmental aspect of the oxygen and hydrogen production. The hybrid system included gasification, organic Rankine cycle (ORC), solid oxide fuel cell (SOFC) and electrolyser sub-systems. The plant achieved hydrogen and oxygen production rates of 67 kg/h and 519 kg/h, respectively, with exergy and energy efficiencies of 84% and 72%, respectively, and an electrical output of 1.1 MW and 52 kW from the SOFC and ORC, respectively. The total capital and operation costs of the system was estimated at 18,204,091$, with an annual income of 3,730.309$. The economic analysis yielded the net present value and payback period of 10,286,826$ and 7.5 years, respectively.

1. Introduction

Medical waste (MW) constitutes a critical environmental and health concern due to its hazardous nature. The proportion of hazardous MW within total hospital waste varies between 15% and 35%,

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depending on the patterns of waste generation [1]. Hospitals are the primary producers of MW, generating larger quantities compared to other facilities such as clinics, laboratories, and autopsy centres [2]. Mismanagement of MW poses serious risks to both the environment and public health, yet it simultaneously presents potential as an alternative energy source [3].

In Jordan, the energy sector is highly vulnerable, as the country imports 96% of its energy while producing only 4% domestically, resulting in an annual import bill of approximately $6.48 billion [4]. Hospitals alone consume about 887,730 kWh monthly at a cost of $227,927.62 [5]. The MW generated in Jordan reached 373,934 tons in 2020 [6], with hazardous fractions ranging from 10% to 25% [7]. This growing volume intensifies environmental and health hazards.

Converting MW into energy is increasingly recognized as a vital strategy for sustainable development. Erdogan and Yilmazoglu [1] emphasized the superiority of gasification, particularly plasma gasification, over traditional incineration methods, reporting hydrogen yields of up to 32% from MW. Plasma gasification is regarded as a promising pathway for hydrogen production due to its high efficiency, low emissions, and ability to handle complex waste streams [8-10]. Several investigations further highlight its economic feasibility, with reported payback periods ranging between 3.5 and 6 years and positive net present values in integrated energy systems [11-13].

Hydrogen, although not naturally abundant, is considered a strategic energy carrier due to its high energy density, carbon-free characteristics, and versatility in storage and utilization [14-16]. Recent studies on MW gasification have demonstrated hydrogen production rates of 38–55% under optimized operating conditions [17,18]. In addition, hybrid system configurations integrating MW gasification with Solid Oxide Fuel Cells (SOFC), gas turbines, or supercritical CO_2 cycles have achieved energy efficiencies exceeding 50% and significant economic returns [11,19].

The urgency of exploring such solutions in Jordan was reinforced by the fatal Al-Salt Hospital accident in March 2021, caused by an oxygen shortage that led to nine deaths [20]. This highlights the critical need for decentralized, reliable, and sustainable oxygen production. The present study proposes a hybrid closed-loop system that integrates MW gasification with Organic Rankine Cycle (ORC) technology to simultaneously generate electricity, hydrogen, and medical oxygen. This system seeks to provide a self-sustained energy supply for hospitals in Amman, while reducing environmental impacts and alleviating the financial burden associated with fossil fuel imports. Moreover, the research aligns with Jordan's national sustainability agenda, including SDGs 7 and 13, Jordan Vision 2025, and the National Climate Change Policy, all of which emphasize renewable energy integration, reduced landfill dependency, and sustainable medical waste management [21-23].

2. Methodology

To establish a numerical framework for the production of hydrogen and oxygen from medical waste (MW) in hospitals located in Amman, the proposed model integrates waste-to-energy technologies with renewable energy sources to achieve sustainable and efficient production pathways. The simulation of the process was conducted using Aspen Plus software, which offers substantial modelling flexibility, extensive databases of physical and thermodynamic properties, and robust modules for chemical reactions and separation processes relevant to hydrogen and oxygen generation.

In this investigation, MW generated in Jordan's capital city is employed as the feedstock, with availability of 220 kg/h, as reported in the statistical records of the Jordanian Ministry of Health, where the flow rates, proximate and ultimate analyses for the MW used in the Aspen Plus simulation were published in our previous work [24]. The model developed using Aspen Plus software is structured into key sub-systems, including gasification, Organic Rankine Cycle (ORC), SOFC and electrolyser as shown in Figure 1.

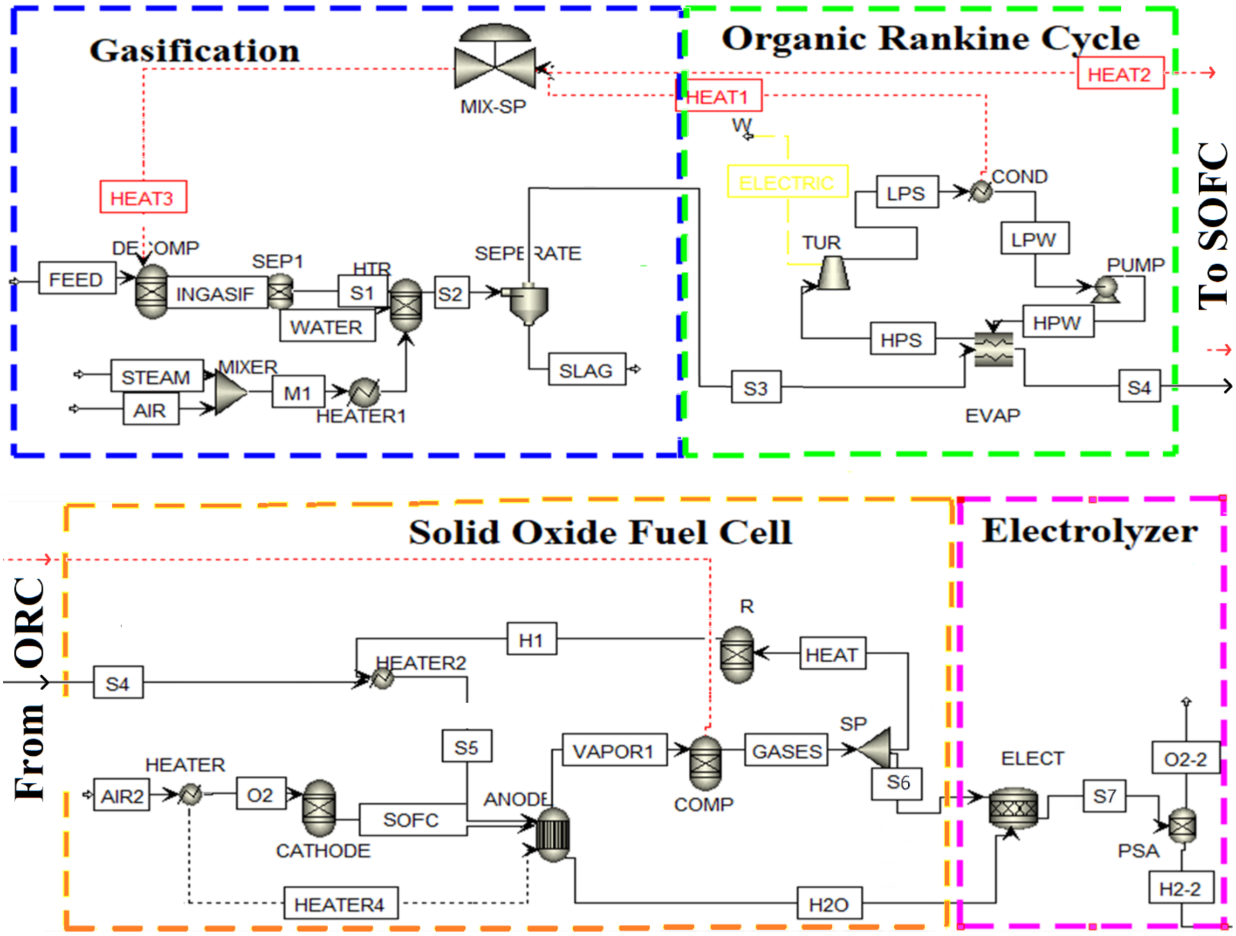

Fig. 1. Hydrogen and oxygen production system layout in Aspen Plus

The central component of the proposed model is the gasification system, which operates with MW, steam, and air as input streams. Although plasma gasification has been highlighted in the literature for its high efficiency and reduced emissions, as discussed in the introduction, its current development remains at an early research and development stage, with experimental applications limited to laboratory-scale trials. Moreover, plasma gasifiers are hindered by high system complexity, elevated capital and operational costs, and incomplete understanding of plasma characteristics. Consequently, the technology is not yet considered viable for large-scale applications from both technological and economic perspectives [1]. In contrast, conventional circulating fluidized bed gasification is a commercially mature and proven technology for large-scale operations [25]. Accordingly, plasma gasification was not adopted in this study due to its prohibitive cost, infrastructure requirements, and dependency on a stable high-power electricity supply, conditions that are currently infeasible in the hospital context of Jordan. Instead, circulating fluidized bed gasification was selected as the most practical and economically feasible option, while plasma gasification is recognized as a promising candidate for future research once the technology advances and supporting infrastructure becomes more readily available.

Prior to gasification, MW undergoes a separation stage, followed by successive processes of drying, thermal decomposition, and chemical reaction, ultimately producing syngas and residual ash. The main governing reactions are presented in Equations (1–5):

$C + H_2O \rightleftharpoons CO + H_2$ (1)

$C + CO_2 \rightleftharpoons 2CO$ (2)

$CO + H_2O \rightleftharpoons CO_2 + H_2$ (3)

$CH_4 + H_2O \rightleftharpoons CO + 3H_2$ (4)

$C + O_2 \rightleftharpoons CO_2$ (5)

The second major subsystem is the Organic Rankine Cycle (ORC), a thermodynamic cycle used to recover waste heat and generate electrical power. It is integrated with other subsystems, including the Solid Oxide Fuel Cell (SOFC) and the electrolyser, to enhance overall system efficiency. The ORC is composed of four primary components: the pump, turbine, evaporator, and condenser.

The SOFC subsystem functions as the principal electricity generator, directly converting chemical energy into electricity and heat via the electrochemical reaction between hydrogen and oxygen. Furthermore, the system utilizes waste heat and unreacted fuel gases to enable integration with secondary power generation units, such as gas turbines or steam turbines, thereby creating a combined-cycle system for enhanced efficiency. In this setup, syngas from the gasifier and air supplied by a blower are directed to the cathode and anode, where the conversion of chemical energy into electrical energy takes place according to the electrochemical reactions in the cathode shown in Equation 6 and in the anode shown in Equations (6–7), which processes water and carbon dioxide into hydrogen, oxygen, and carbon-based gases.

$$\frac{1}{2}O_2 + 2e^- \rightarrow 0^{-2} \quad (6)$$

$$H_2 + 0^{-2} \rightarrow H_2 0 + 2e^- \quad (7)$$

$$CO + 0^{2-} \rightarrow CO_2 + 2e^- \quad (8)$$

The hydrogen and oxygen products are subsequently separated using adsorption-based technologies. Two techniques are widely discussed in the literature: Temperature Swing Adsorption (TSA) and Pressure Swing Adsorption (PSA). TSA operates through cyclic heating and cooling up to approximately 200 °C, resulting in significant thermal losses and increased exergy destruction. PSA, by contrast, operates at ambient temperature using pressure cycling, which avoids heat losses and achieves lower exergy destruction, despite the energy penalty associated with gas compression [26]. Given these advantages, PSA was selected for gas separation in this study. The PSA unit is available within the Aspen Plus database, where gases, flow rates, and energy requirements are specified directly.

MW in Jordan is characterized by a high volatile matter content (70–80%), comparable to biomass fuels such as wood, empty fruit bunch, and coconut shell [27], but with a relatively lower hydrogen content. This composition makes MW unsuitable for direct combustion due to excessive smoke formation at low temperatures, arising from rapid devolatilization. Consequently, gasification represents the most appropriate conversion pathway to produce syngas. For conventional biomass air-gasification, the equivalence ratio (ER) typically ranges between 0.2 and 0.4. At ER values below 0.2, the process is dominated by pyrolysis, producing primarily charcoal rather than syngas, while ER values above 0.4 promote combustion, yielding excessive CO_2 and non-combustible syngas [17]. Hence, ER values of 0.2, 0.3, and 0.4 were selected for system characterization and optimization.

Due to the inherently low hydrogen content in MW, additional steam was introduced as a gasifying agent to increase H_2 yields. Literature indicates that a minimum steam-to-biomass (S/B) ratio of 1 is required to initiate H_2 production, while ratios exceeding 2 results in incomplete reactions, cooling of the gasifier, and reduced efficiency [28]. Thus, S/B ratios of 1, 1.5, and 2 were considered in this study.

Gasification temperature was also investigated, with reported minimum and maximum values of 600°C and 1000°C, respectively, for activation of reduction-zone reactions in biomass gasification [29]. Accordingly, gasification temperatures of 600°C, 800°C, and 1000°C were selected. The analysis evaluates the influence of ER, S/B ratio, and temperature on syngas composition, hydrogen yield, and overall system performance. Energy and exergy balances were subsequently applied to determine the optimal operating conditions using Equations (8–15):

$$\text{Energy input} = W_{in} + Q_{in} \quad (8)$$

$$\text{Energy output} = W_{out} + Q_{out} \quad (9)$$

$$\text{Total energy} = (W_{in} + Q_{in}) - (W_{out} + Q_{out}) \quad (10)$$

$$\text{Total exergy} = \text{chemical exergy} + \text{physical exergy} \quad (11)$$

$$\text{Chemical exergy} = \Sigma x_i \cdot ex_i \quad (12)$$

$$\text{Physical exergy} = (H - H_0) - T_0 (S - S_0) \quad (13)$$

$$\text{Exergy efficiency}, \eta ex = \frac{\text{useful exergy output}}{\text{total exergy input}} \times 100\% \quad (14)$$

$$\text{Exergy destruction} = \text{Exergy output} - \text{Exergy input} \quad (15)$$

Where x_i is the mole fraction and ex_i is the specific chemical exergy of each component. H = enthalpy, S = entropy, and H_0, S_0 correspond to reference conditions at T_0. To determine the net present value and payback period of hydrogen and oxygen production model through MW gasification. The net present value (NPV) and payback period (PBP) for the optimal energy production case will be calculated based on the model developed in the previous objectives using equations 16 and 17.

$$\text{PBP} = \frac{\text{Initial Investment Cost}}{Annual\ cash\ flow} \quad (16)$$

$$\text{NPV} = \sum \frac{Ct}{(1+r)t} - \text{C} \quad (17)$$

Where: **Ct** = Cash flow in year t; **r** = Discount rate; t = Number of years; **C** = Initial investment cost.

3. Results

The manipulative variables of temperature, ER and S/B ratio have direct effect on gasification process, mainly on syngas quality which is characterized in terms of gas composition and LHV, as well as the sensible heat and flow rate of syngas out of the gasifier. The change in syngas characteristics, on the other hand, have indirect effects on the hybrid system performance in general which is characterized by the performance of the other sub-systems of ORC, SOFC and electrolyser. In order to investigate the effect of the variables on gasification, the quality of the dry and cold syngas after the gasifier at point S4 is analysed in terms of syngas composition and LHV. This is followed by the chemical and electrical cogeneration capacity and the overall system energy and exergy analysis.

3.1 Impact of Gasification Temperatures on System Performance

Water and particles are separated from syngas after the gasifier and then the sensible heat of syngas is recovered to provide the needed heat for the ORC sub-system for electrical generation. Figure 2 illustrates the volume composition and LHV model results for cold and dry syngas at point S4 at the exit of ORC sub-system. The analysis considers three gasification temperatures of 600°C, 800°C and 1000°C, while maintaining ER and S/B ratio values as constant at their maximum values of 0.4 and 2, respectively. Gasification temperature plays a critical role in governing the reaction pathways within the gasifier, as several reactions are constrained by activation energy barriers that are strongly temperature dependent.

In the present system, which employs both air and steam as gasification agents, significant CO_2 formation occurs near the air inlet due to direct fuel oxidation. The reversed Boudouard reaction (Eq. 2), which reduces CO_2 to CO, becomes particularly important in counteracting this complete oxidation. However, this reaction requires relatively high activation energy, with a threshold of approximately 710°C reported for initiation in the absence of a catalyst [30]. This mechanism explains the sharp decline in CO_2 concentration from 19% at 600°C, where the Boudouard reaction is largely inactive, to 12% at 1000°C, where it becomes dominant. Consequently, CO levels increase markedly from 7% to 16% as the temperature rises. At lower temperatures, CO formation is primarily attributed to partial carbon oxidation, while steam reforming reactions remain minimally active, thus contributing little to CO production at 600°C.

Methane reforming is another high-activation-energy reaction that converts CH_4 into CO and H_2 in the presence of steam. Its activation with rising temperature accounts for the observed reduction in CH_4 content dropping below 1%, alongside modest increases in both H_2 and CO. The increase in syngas LHV with temperature was relatively limited, since the gain in CO was offset by concurrent decreases in H_2 and CH_4 fractions.

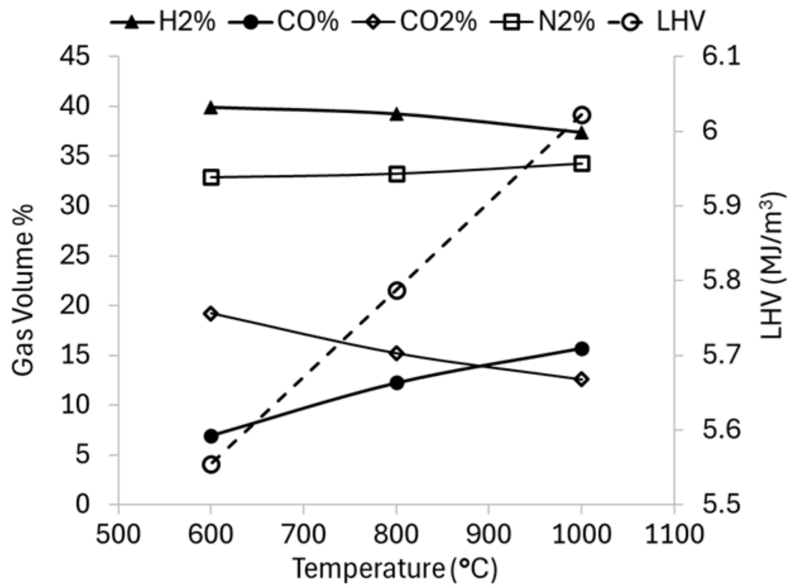

Fig. 2. The impact of temperatures on syngas volume composition and LHV

Gasification temperature exerts an indirect influence on the downstream subsystems—namely the ORC, SOFC, and electrolyser—as well as on the chemical production of H_2 and O_2 and the overall hybrid system performance. This occurs primarily through its direct impact on syngas composition and its sensible heat, which is subsequently utilized to supply the thermal input required by the ORC. The results indicate that gasification temperature has only a marginal effect on hydrogen production, while its influence on oxygen generation is more pronounced. Both hydrogen and oxygen outputs increase proportionally with temperature, with maximum yields achieved at 1000 °C, as shown in Table 1. The electrolyser outlet stream (point S7) consists mainly of H_2, O_2, and CO_2, with minor traces of CH_4 (<1 kg/h).

The hybrid system incorporates five subsystems, of which the SOFC and ORC serve as the main power generation units. Simulation results reveal that the SOFC attained its maximum electrical output at 600°C, generating 1142 kW, while the minimum output of 1084 kW was recorded at 1000°C. This decline is attributed to the reduction in H_2 yield from the gasifier at elevated temperatures, caused by suppression of steam reforming reactions by the Boudouard reaction. In contrast, the ORC exhibited an increasing output with higher gasification temperatures, benefiting from the additional sensible heat available in the syngas. The ORC output rose from approximately 51.1 kW at 600°C to nearly 51.7 kW at 800–1000°C. The system's energy and exergy efficiencies were evaluated across different gasifier temperatures. The highest energy efficiency was observed at 800°C, reaching 72.66%, followed by 69.74% at 1000°C. The lowest energy efficiency occurred at 600°C, where the system achieved 64.25%. In contrast, the highest exergy efficiency was observed at 600°C, reaching 84.08%, which is primarily attributed to the relatively low exergy destruction of 100.3 kW at this condition. The second-highest efficiency was obtained at 800°C, with a value of 83.32%. These results indicate a strong dependence of exergy efficiency on the magnitude of exergy destruction. Interestingly, the exergy destruction at 1000°C (113.21 kW) was higher than that recorded at 600°C and 800°C, as presented in Table 1. This behaviour reflects an inverse relationship between gasification temperature and system exergy efficiency, suggesting that although higher temperatures can reduce certain irreversibility within the gasifier, they may also introduce additional losses that diminish overall efficiency. This also accounts for the slight reduction in energy efficiency observed at 1000 °C, where elevated operating conditions contribute to higher system losses.

Table 1
System outputs at three different temperatures levels

T (°C)	H_2 (kg/h)	O_2 (kg/h)	CO_2 (kg/h)	CH_4 (kg/h)	SOFC (kW)	ORC (kW)	Energy efficiency (%)	Exergy efficiency (%)	Exergy destruction (kW)
600	67.38	518.20	125.37	0.445	1142.32	51.06	64.25	84.08	100.3
800	67.39	519.26	124.005	0.405	1105.14	51.69	72.66	83.32	106.6
1000	67.40	520.35	122.442	0.364	1084.05	51.63	69.74	82.53	113.2

3.2 Impact of Equivalent Ratio on System Performance

The second investigated variable is the equivalence ratio (ER), which was varied from 0.2—representing low air input and pyrolysis-dominated air-gasification—up to 0.4, where higher air input results in combustion-dominated gasification. Figure 3 illustrates the effect of increasing the air inlet flow on syngas composition and its LHV at a fixed gasification temperature of 1000°C and an S/B ratio of 2. The results show that increasing air supply leads to greater dilution of syngas with nitrogen, rising from 19% to 34%. This dilution reduces the relative fractions of other major constituents, particularly H_2 and CO, which in turn lowers the LHV from 7.8 MJ/m^3 to 6 MJ/m^3. Furthermore, at 1000°C the Boudouard reaction remains dominant, contributing to the reduction in CO_2 concentration within the syngas.

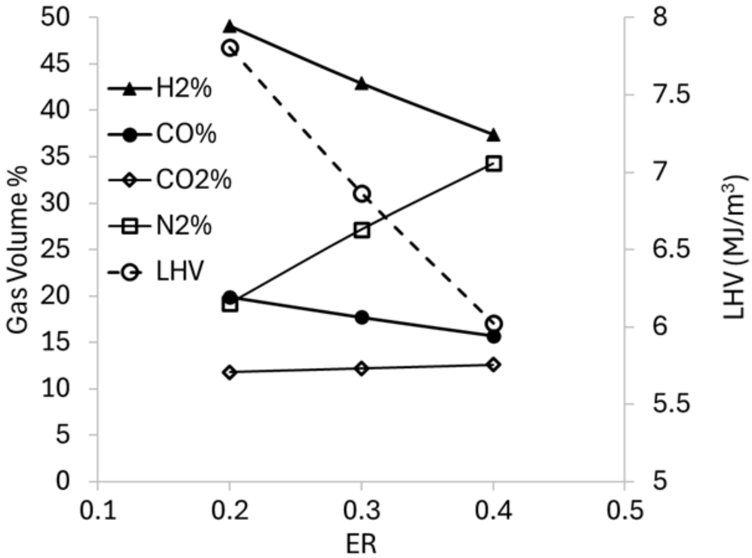

Fig. 3. The impact of ER on syngas volume composition and LHV

The influence of varying ER in the range of 0.2–0.4 was further examined with respect to hydrogen and oxygen production rates, the LHV of the product gas stream at point S7, and overall system performance in terms of electrical output, efficiency, and exergy destruction, as shown in Table 2. With the fuel flow rate fixed at 220 kg/h, increasing ER corresponds to higher air input, which increases the overall syngas mass flow. This results in higher sensible heat and elevated oxygen content in the syngas due to the additional air supply to the gasifier.

The results show that hydrogen production reached a maximum of 67.40 kg/h at ER of 0.4, compared to 67.29 kg/h at ER of 0.2, indicating only a marginal dependence on ER. Oxygen production followed a similar trend, with the highest output of 520.35 kg/h observed at ER of 0.4, compared to 515.79 kg/h at ER of 0.2. The LHV, however, demonstrated an inverse correlation with ER. In terms of system performance, increasing ER enhanced the electrical output of both the SOFC and ORC subsystems. At ER of 0.4, the SOFC produced 1084.05 kW and the ORC 51.63 kW, whereas the lowest outputs were obtained at ER of 0.2, with 977.35 kW from the SOFC and 45.71 kW from the ORC. These improvements can be attributed to the higher syngas flow and associated increase in sensible heat, as well as the greater availability of exhaust heat recovered for the ORC and SOFC.

The system analysis indicated that the maximum energy efficiency was achieved at an ER of 0.3, with a value of 70.78%, while the lowest efficiency of 69.42% occurred at an ER of 0.2. Furthermore, a direct correlation was observed between ER and exergy efficiency. The highest exergy efficiency, 82.53%, was recorded at ER of 0.4, which corresponded to the lowest exergy destruction among all studied cases of 113.21 kW. By contrast, exergy destruction at ER of 0.2 reached 191.75 kW, while at ER of 0.3 it declined to 152.50 kW. These results reinforce the inverse relationship between ER and exergy destruction, underscoring the role of higher air input in minimizing irreversibility within the system.

Table 2
System outputs at three equivalent ratios

ER	H_2 (kg/h)	O_2 (kg/h)	CO_2 (kg/h)	CH_4 (kg/h)	SOFC (kW)	ORC (kW)	Energy efficiency (%)	Exergy efficiency (%)	Exergy destruction (kW)
0.2	67.29	515.79	56.85	0.77	977.35	45.71	69.42	73.08	191.75
0.3	67.36	518.86	88.57	0.50	1030.40	48.67	70.78	77.80	152.50
0.4	67.40	520.35	122.44	0.36	1084.05	51.63	69.74	82.53	113.21

3.3 Impact of S/B Ratio on System Performance

The final investigated variable is the steam-to-biomass (S/B) ratio, which was varied to examine the influence of steam on the gasification process. Since the Boudouard reaction becomes dominant at elevated temperatures, the operating conditions were set to 800°C and an ER of 0.3 to limit the activation of this reaction. This approach reduces the competition among solid–gas reactions and enhances the likelihood of activating steam reforming reactions of both carbon and methane. The resulting dry syngas composition and corresponding LHV at point S4 are presented in Figure 4.

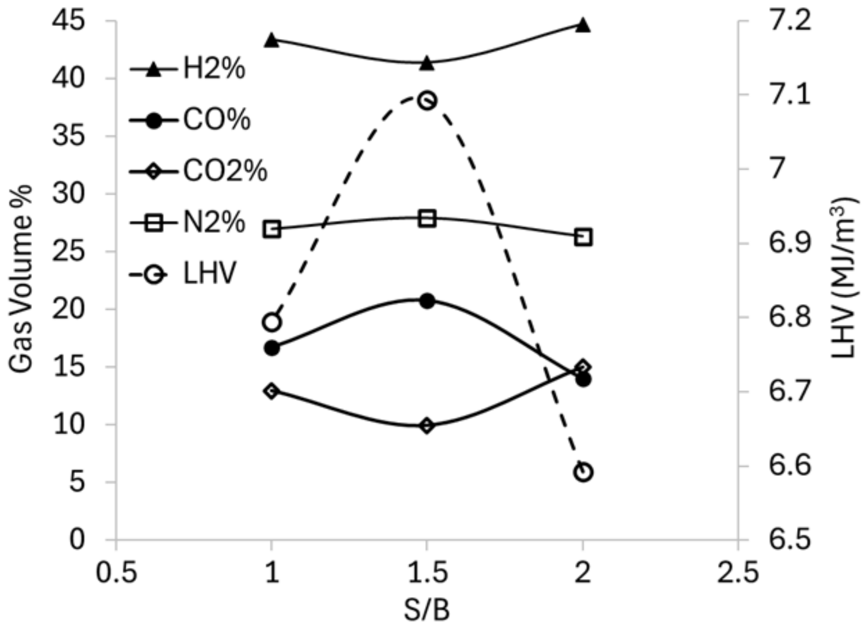

Fig. 4. The impact of S/B on syngas volume composition and LHV

Table 3 presents the effect of S/B ratios on the overall hybrid system in terms of the outputs for hydrogen and oxygen production, electricity output from the ORC and SOFC, energy efficiency, exergy efficiency and exergy destruction. These results correspond to three S/B ratios of 0.1, 1.5, and 2, while maintaining constant values for both gasifier temperature and ER. To characterize the effect of steam flow on system performance, the equivalence ratio and gasification temperature were fixed at 0.3 and 800°C, respectively, to suppress the Boudouard reaction and allow steam reforming reactions of carbon and methane to fully activate. This approach increased the H_2 fraction in the syngas and influenced the CO_2 content in the final product stream at S7. Figure 4a presents the hydrogen and oxygen production rates across three S/B ratios. The results indicate that increasing the steam input to an S/B ratio of 2.0 produced the highest yields, with approximately 67.35 kg/h of hydrogen and 517.66 kg/h of oxygen. In contrast, an S/B ratio of 1.0 resulted in lower production, generating 42.77 kg/h of hydrogen and 325.81 kg/h of oxygen.

System electrical outputs also improved with higher steam input. At an S/B ratio of 2.0, the SOFC and ORC generated 1051.54 kW and 48.73 kW, respectively. For S/B of 1.5, the output decreased to 839.51 kW (SOFC) and 43.30 kW (ORC), while at S/B of 1.0, they further declined to 621.46 kW (SOFC) and 37.81 kW (ORC). Interestingly, energy efficiency exhibited an inverse trend with S/B ratio. The highest energy efficiency of 75.17% was observed at S/B of 1.0, whereas the lowest, 69.4%, occurred at S/B of 2.0, primarily due to the higher energy demand associated with producing increased steam flow. Conversely, exergy efficiency improved with higher S/B ratios, attaining a maximum of 78.57% at S/B of 2.0. This corresponds to a lower exergy destruction of 146.05 kW at this ratio, compared to 200.43 kW at S/B of 1.5 and 254.50 kW at S/B of 1.0, as shown in Table 3. These findings reinforce the observed inverse relationship between exergy destruction and S/B ratio.

Table 3
System outputs at three steams to biomass ratios

S/B	H_2 (kg/h)	O_2 (kg/h)	CO_2 (kg/h)	CH_4 (kg/h)	SOFC (kW)	ORC (kW)	Energy efficiency (%)	Exergy efficiency (%)	Exergy destruction (kW)
1	42.77	325.81	85.47	0.398	621.46	37.81	75.17	64.25	254.50
1.5	55.05	420.86	88.97	0.509	839.51	43.30	72.10	71.55	200.43
2	67.35	517.66	90.15	0.549	1051.55	48.74	69.40	78.57	146.05

3.4 The Economic Analysis for the Model

The findings indicate that best performance was achieved at gasification temperature of 800°C, S/B 2.0, and ER of 0.4, which yielded hydrogen and oxygen production rates of 67 kg/h and 519 kg/h, respectively. Therefore, the economic analysis was performed at these operating conditions. The total cost of the gasification process was estimated at $11,424,340, as presented in Table 4. Based on the literature, the capital cost of the ORC system is 2000 €/kW [31], while the SOFC capital cost is reported as $1,848,220 [32], as shown in Table 4. The capital cost of the composition chamber is $61,230 [33].For the electrolyser, the capital cost is 388 €/kW [34], and for the PSA unit, it amounts to $1,841.02 [32]. Summing these values, the total capital cost of the integrated system is $16,549,174. It must be noted, the operation cost is 1,654917.4 as illustrated in Table 5, the lifetime of economic period is 18 years.

Table 4
Model cost summary

Model Items		**Cost**	**References**
Gasifier	Gasifier	6,080,970 $	[32]
	ASU	5,314,010 $	
	H1	14,680 $	
	H2	14,680 $	
	H3	14,680 $	
SOFC	$_A$SOFC	1,532,910 $	
	$_Z$SOFC	153,290 $	
	AB	86,390 $	
	HE	19,680 $	
	CC	30,400 $	
	DC	25,530 $	
ORC	ORC	2000 €/kWh	[31]
Composition Chamber	Composition Chamber	61,230 $	[33]
Electrolyser	Alkaline electrolysis	388€/kW	[34]
PSA	PSA	1,841,020 $	[32]
Capital Cost		16,549,174$	

Table 5
Basic assumptions for economic analysis

Item	Unit	Value	References
Annual operation time	h	7000	
Lifetime of WTE system (Economic Period)	Year	18	[35]
Maintenance and operation cost	$	10% of capital cost	
Discount rate	%	10	[36]
Inflation	%	1.98	[37]

The economic analysis indicated that the annual revenue from electricity production was approximately $908,797, while revenue from hydrogen sales amounted to $1,922,213, based on the sale of 589 tons, as summarized in Table 6. The project achieved a net present value (NPV) of $10,286,826 and PbP of 7.50 years, as illustrated in Figure 5.

Table 6
The model income summary

Item	Value
Annual amount of treated medical waste (t)	1929
Annual production of electricity (MW)	10083.41
Annual production of H2 (t)	589.920
Annual income due to waste treated (K$)	889.2998
Annual income due to electricity (K$)	908.797
Annual income due to hydrogen sales (K$)	1922.213
Total income (K$)	3730.309
NPV (K$)	10,286.826
Payback Period	7.50 Years

This illustrations of the annual cash flows of the project show the difference between the cash flows involved and out of the life of the project. In Figure 5 The orange column represents the net cash flow for each year, while the green column shows cumulative cash flow. The blue line also represents cumulative cash flows in line with the Payback Period. Through this drawing, the performance of the financial project can be evaluated, monitoring profit growth, and setting the time required to recover initial investment.

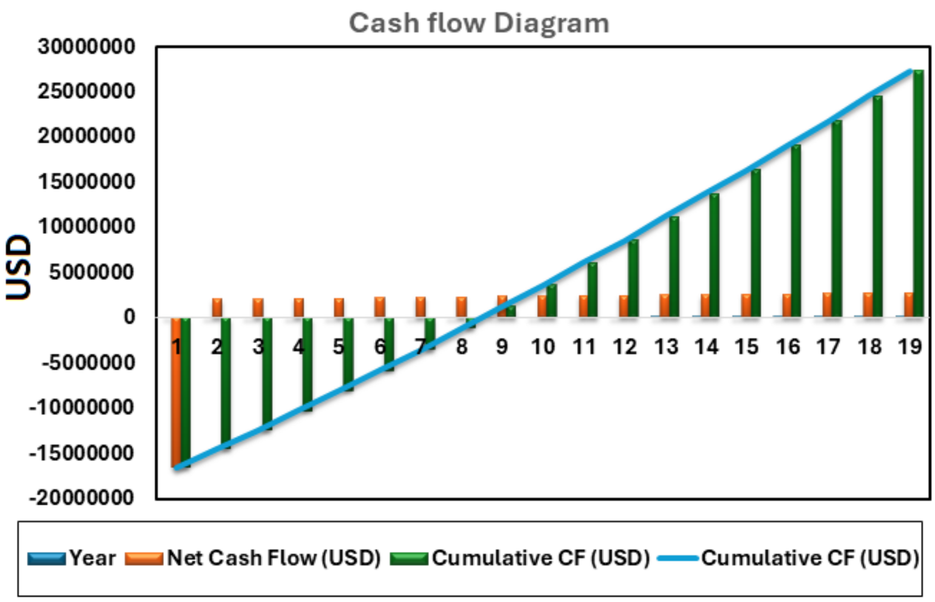

Fig. 5. Cash flow diagram

4. Conclusions

This study aims to develop a model for the conversion of MW into hydrogen and oxygen, specifically for application in hospitals located in Amman, the capital of Jordan. The research also investigates the influence of key operating parameters—namely gasification temperature, ER, and S/B —on model performance. and the local market capacity in Jordan for the utilization of the produced hydrogen, and the economic feasibility of the optimum hybrid system design based of the equipment, labour, maintenance costs and electric tariff commercial rates in Jordan. The best operating condition were achieved at temperature of 800°C, Er of 0.4 and S/B ratio of 2.0. The system can generate a total electrical output of 1.15 MW from the SOFC and ORC that enables self-sustained operation fuelled by medical waste in Amman city while providing a solution for the disposal of the waste that is mostly disposed in the landfills of the city. The developed model demonstrates the system's significant potential for economic analysis, with annual revenues generated from electricity production estimated at $908,797, while total annual revenues from hydrogen sales amounted to $1,922,213, derived from the sale of 589 tons. The project achieved NPV of $10,286,826 and PbP of 7.50 years, making it an attractive investment for central plant hydrogen and oxygen generation for Amman city.

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