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# Development of the Conical Spouted Bed Technology for Biomass and Waste Plastic Gasification

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## Abstract

Gasification is one of the most effective methods for upgrading different wastes, such as plastics and biomass, because the gas produced can be used directly as a fuel or as a renewable raw material for the production of chemicals and fuels. The conical spouted bed reactor (CSBR) has demonstrated to perform well in gasification process due to its specific features, such as (i) the cyclic and vigorous particle movement that avoids bed defluidization (a limitation in fluidized beds), (ii) capability for handling irregular or sticky solids, (iii) high heat transfer rates between phases, and (iv) bed stability in a wide range of gas flow rates. However, the conventional CSBR is characterized by its short residence time, which involves serious problems for minimizing tar formation. The incorporation of a fountain confiner in the CSBR is key to increasing the gas residence time and improving the contact between the gas and heat carrier particles, thereby promoting tar cracking reactions and so enhancing carbon conversion efficiency from 81.5% (without confiner) to 86.1% under fountain enhanced regime. The quality of the syngas is clearly improved as the  $H_2$  concentration increases from 36 to 42% with and without the fountain confiner, whereas that of CO decreases from 34 to 29%, respectively.

**Keywords:** gasification, conical spouted bed, fountain confinement, syngas, biomass, plastic waste

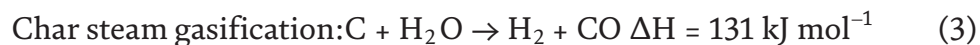
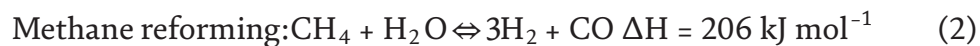
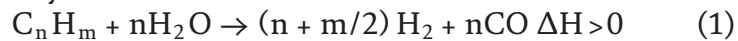
## 1. Introduction

Gasification is a thermochemical process that transforms carbonaceous materials (coal, oil and its derivatives, biomass, post-consumer and industrial solid wastes) into syngas, with CO and  $H_2$  being its major components. The gasification process takes place at high temperatures (generally in the 600–900°C range or even higher) in the presence of a gasifying agent (air, oxygen, steam,  $CO_2$ , or mixtures of these components) at a lower ratio than that stoichiometrically required for combustion. Syngas production is essential due to the increasing interest in gas to liquid (GTL) processes through the synthesis of methanol, dimethyl ether, and Fischer-Tropsch. In addition, the valorization of syngas can be integrated with

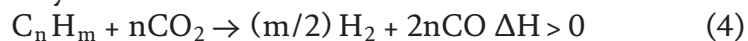
energy recovery systems, by means of turbines, combined cycle units, or fuel cells. The gasification technology has been extensively developed for coal and oil products and is gaining increasing interest for biomass [1, 2] in which catalysts play an essential role [3]. Furthermore, the upgrading of post-consumer solid wastes by gasification is becoming a short-term promising strategy [4].

Gasification involves several steps and complex chemical reactions, which may be grouped as follows: drying, pyrolysis, cracking and reforming reactions in the gas phase, and heterogeneous char gasification. The significance of these steps on the process performance and their kinetics depends on the feedstock characteristics and gasification conditions. The pyrolysis step involves a series of complex chemical reactions of endothermic nature and leads to volatiles (gases and tars) and a solid residue or char. The homogeneous gasification reactions include a wide variety of reactions, with the balance and the extent of these reactions depending mainly on the gasifying agent used, its ratio with respect to the feed (S/feed ratio), and temperature. These reactions are as follows:

Steam reforming of hydrocarbons:



Dry reforming of hydrocarbons:



It should be noted that gasification reactions are only those involving  $H_2O$  and  $CO_2$ , because  $O_2$  only promotes combustion and partial oxidation reactions that produce  $CO$ ,  $CO_2$ , and  $H_2O$ . In addition, the exothermic nature of oxidation reactions provides the energy required for the highly endothermic steam and  $CO_2$  reforming (Eqs. (1)–(4)) and Boudouard (Eq. (5)) reactions. Steam improves  $H_2$  production by means of steam reforming reactions (Eqs. (1) and (2)) and also by enhancing the water-gas shift (Eq. (6)) equilibrium. High temperatures are required for promoting char gasification, especially  $CO_2$  gasification, whose kinetics is between 2 and 5 times slower than under steam atmosphere and does not occur below  $730^\circ\text{C}$  [5].

The main drawback of the syngas produced is the presence of certain impurities, such as fine particles, organic tars,  $NO_x$ , and  $SO_2$ , which need to be removed before its application in subsequent processes [6]. In particular, tar is the main contaminant in the gas produced, and its content ranges from 5 to  $100 \text{ g Nm}^{-3}$ , depending on the type of gasifier. However, its maximum allowable content is  $5 \text{ mg Nm}^{-3}$  in gas turbines and  $100 \text{ mg Nm}^{-3}$  in internal combustion engines [7, 8]. Tar is described as a complex mixture of condensable hydrocarbons, ranging from single-ring to

five-ring aromatic compounds along with other oxygen-containing hydrocarbons and complex polycyclic aromatic hydrocarbons (PAHs) [9]. These compounds may cause several operational problems, such as condensation and the subsequent plugging of downstream equipment, clogging filters, and metal corrosion, which lead to unacceptable levels of maintenance for engines and turbines.

All the methods available for tar reduction may be classified into two groups, depending on where tar is removed: in situ (or primary) methods and post-gasification (or secondary) methods. Regardless of the strategy followed, the optimum operating conditions, appropriate additives or catalysts, and a suitable reactor configuration should be established in order to obtain a gas stream with a maximum tar content of  $2 \text{ g Nm}^{-3}$  and a low content of PAH compounds [10, 11]. It should be noted that tar formation depends on the gasification conditions, particularly on temperature, so preventive treatments are recommended to operate above  $1000^\circ\text{C}$ . Aznar et al. [12] suggest injecting a secondary air stream into the freeboard to reduce the content of tar.

In situ catalytic cracking is one of the most promising techniques, as it allows reducing the need for expensive downstream operations [3, 13]. Natural minerals, such as olivine [14, 15] and dolomite [16], have been widely used in steam gasification because, apart from being active for the cracking and reforming of heavy aromatic compounds, they are inexpensive and abundant. In addition, Ni catalysts have received great attention in gasification due to their higher effectiveness for converting tar into  $\text{H}_2$ -rich gas [17, 18].

Moreover, apart from temperature and catalysts, reactor design also plays a critical role in gasification. Different reactor configurations are commonly used for the steam gasification process, which according to their hydrodynamic behavior can be classified as follows: fixed bed, fluidized bed, entrained flow, and rotary kiln reactors, among others [4]. Fluidized beds are the most commonly used due to their advantages, such as versatility for using different types of wastes (agroforestry, post-consumer, and industrial), high heat and mass transfer rates between phases, and bed isothermicity, which allow the scaling-up of the process to the industrial level [19–21]. Nevertheless, biomass or waste particles of irregular texture require a large amount of inert solid (sand) to promote their fluidization. In addition, small particle sizes (Geldart A and B) are the best for fluidization, and therefore high amounts of energy are required to grind and sieve the feedstock. Nevertheless, there is an alternative to conventional fluidized beds, namely, the conical spouted bed reactor (CSBR), which may handle residues of different densities and sizes without significant segregation in the bed. This technology allows handling larger particles than those in fluidized beds, including those with an irregular texture, fine materials, and sticky solids, with no agglomeration or segregation problems [22]. Moreover, the highly vigorous movements of the solids lead to high heat and mass transfer rates between phases [23]. Other advantages of the CSBR over the fluidized bed are its simpler design (no distributor plate) and the lower sand/feed ratio required for the same capacity.

The main drawback of this technology for gasification is the short gas residence time, which hinders tar cracking reactions. Accordingly, certain modifications have been developed in order to improve its performance in the gasification process by changing reactor hydrodynamics, which are as follows: the confinement of the fountain and the use of draft tubes. The fountain confinement device is a tube welded to the lid of the reactor that allows operating under stable conditions with fine particles and increasing the gas residence time by lengthening the path followed by the gas [24]. Therefore, gas-solid (catalyst) contact in the fountain is greatly improved, and tar cracking and reforming reactions are therefore promoted. Moreover, the draft tube also enables to widen

the operation range and improve the reactor's hydrodynamic behavior [24]. Thus, this chapter summarizes the main results obtained in the application of the conical spouted bed reactor in the steam gasification of biomass and waste plastics. Moreover, the influence of different primary catalysts and the incorporation of novel modifications in the reactor design, such as fountain confiner and draft tube, are also discussed.

## 2. Material and methods

### 2.1 Feedstock characterization

The HDPE was supplied by Dow Chemical (Tarragona, Spain) in the form of chippings (4 mm), with the following properties: average molecular weight,  $46.2 \text{ kg mol}^{-1}$ ; polydispersity, 2.89; and density,  $940 \text{ kg m}^{-3}$ .

The biomass used in this study is forest pinewood waste (*Pinus insignis*). The sawdust has been sieved to obtain a particle size between 1 and 2 mm. This material has been dried at room temperature to a moisture content below 10 wt%. Ultimate and proximate analyses have been carried out in a LECO CHNS-932 elemental analyzer and in a TGA Q500IR thermogravimetric analyzer, respectively. The high heating value (HHV) for both biomass and HDPE was measured in a Parr 1356 isoperibolic bomb calorimeter. The main features of both the raw biomass and the HDPE are summarized in **Table 1**.

### 2.2 Equipment

Steam gasification runs have been carried out in a bench-scale plant, whose scheme is shown in **Figure 1**. The main element of the plant is the conical spouted bed reactor (CSBR), whose design is based on previous hydrodynamic studies [25] and on the application of this technology to the pyrolysis of different solid wastes, such as biomass [26–28], plastics [29], and waste tires [30].

	Biomass	HDPE
Ultimate analysis (wt%)		
Carbon	49.33	85.71
Hydrogen	6.06	14.29
Nitrogen	0.04	—
Oxygen	44.57	—
Proximate analysis (wt%)		
Volatile matter	73.4	99.7
Fixed carbon	16.7	0.3
Ash	0.5	—
Moisture	9.4	—
HHV ( $\text{MJ kg}^{-1}$ )	19.8	43.1

**Table 1.**  
Characterization of the biomass and HDPE used in this study.

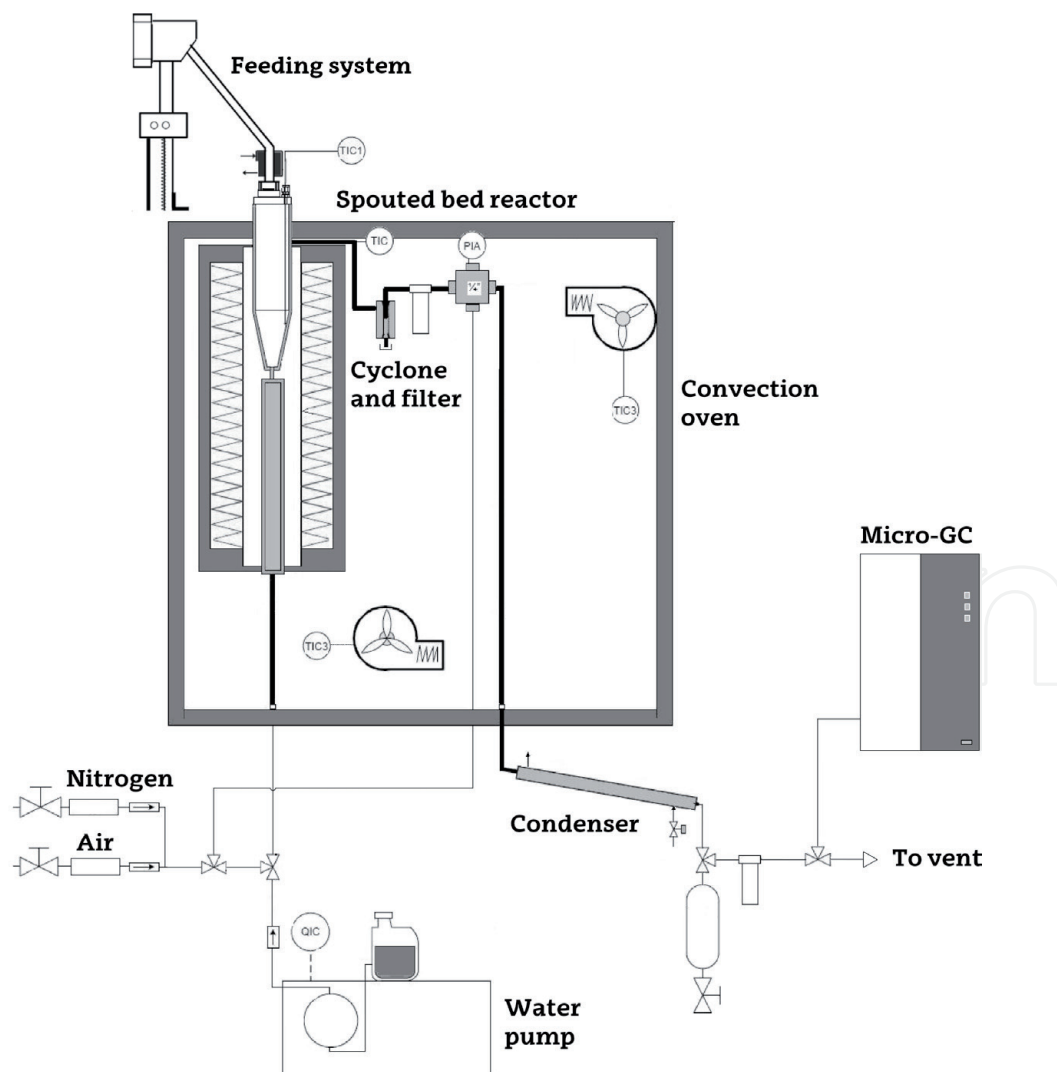


The plant is provided with a system for continuous feeding of the biomass or plastic. The system for solid feeding consists of a vessel equipped with a vertical shaft connected to a piston placed below the material bed. The plastic/sawdust is fed into the reactor by raising the piston at the same time as the whole system is vibrated by an electric engine.

Water has been fed by means of a Gibson 307 pump that allows a precise measuring of the flow rate. The water stream has been vaporized by means of an electric cartridge placed inside the forced convection oven and prior to the entrance of the reactor.

The reactor is located within an oven, which is in turn placed in a forced convection oven maintained at 270°C to avoid the condensation of steam and tars before the condensation system. A high-efficiency cyclone and a sintered steel filter (5 µm) are also placed inside this oven in order to retain the fine sand particles entrained from the bed and the soot or char particles formed in the gasification process.

The gases leaving the forced convection oven circulate through a volatile condensation system consisting of a condenser, a Peltier cooler, and a coalescence filter. The Peltier cooler consists of a 150 mL tank and a refrigerator that lowers the temperature to around 2°C, thereby efficiently condensing the volatile products. The condenser is a double-shell tube cooled by tap water.



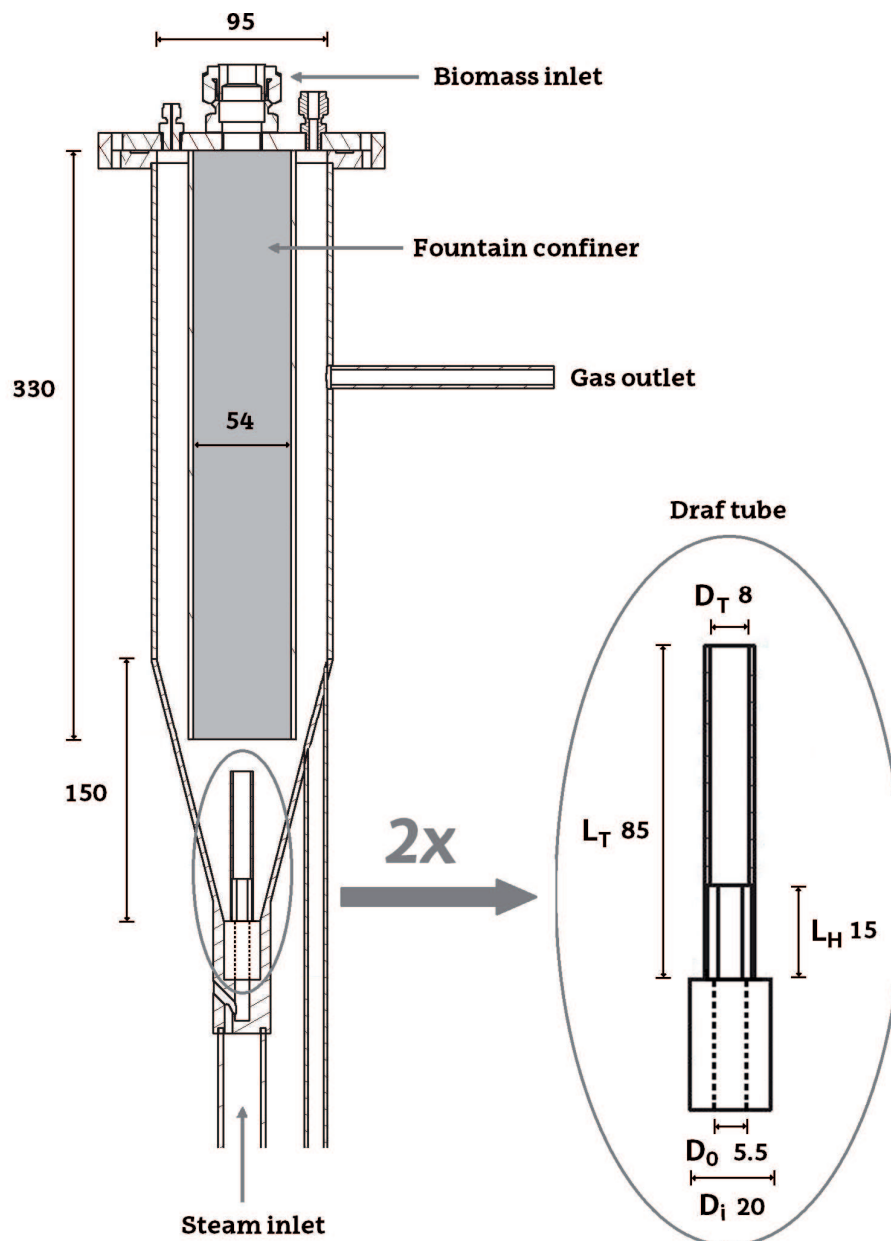
**Figure 1.**  
*Scheme of the bench-scale biomass gasification plant equipped with a conical spouted bed reactor.*

### 2.2.1 Conventional spouted bed reactor

The spouted bed reactor is the core of the gasification plant. The total height of the reactor is 298 mm, with that of the conical section (angle of  $30^\circ$ ) being 73 mm. The diameters of the cone base and cylindrical section are 12.5 and 60.3 mm, respectively. The gas inlet diameter is 7.6 mm. Despite the endothermic nature of the steam gasification process, bed isothermicity is ensured by the vigorous solid circulation of the sand in this reactor, which also promotes high heat transfer rates [23]. The CSBR is placed inside a 1250 W radiant oven. Two K-type thermocouples are located inside the reactor, one in the bed annulus and the other one close to the wall.

### 2.2.2 Fountain-enhanced spouted bed reactor

This reactor is an improved version of that described in Section 2.2.1, which has been specifically designed for gasification process. Thus, a fountain confiner was welded to the lid in order to increase the residence time, narrow its distribution, and improve the gas-solid contact in the fountain region (**Figure 2**). Thus, several



**Figure 2.** Main dimensions (in mm) of the spouted bed gasifier, fountain confiner, and draft tube.

modifications were introduced in order to optimize its performance. For example, the height of the reactor was increased in order to increase the residence time of the gas and promote tar cracking. This reactor may also operate in the conventional spouting regime by using a lid without confiner. It is noteworthy that its design allows using draft tubes to widen the application range of the spouting regime and improve bed stability [31, 32]. In fact, the nonporous draft tube promotes high fountains [32] by diverting most of the inlet gas stream through the draft tube, which also enhances solid cross-flow from the annulus into the spout and therefore leads to additional gas-solid contact in the fountain.

The main dimensions of this spouted bed reactor, the fountain confiner, and the draft tube used are depicted in **Figure 2**. According to a previous hydrodynamic study conducted under gasification conditions [33], a draft tube with 8 mm in external diameter (5.5 mm in internal diameter) and 15 mm entrainment zone height was determined as the optimum one. Thus, these geometric factors allow operating under enhanced fountain regime, with low steam flow rates ensuring great turbulence and a well-developed fountain region with a great hydrodynamic stability.

### 2.3 Primary catalysts

$\gamma$ -Al<sub>2</sub>O<sub>3</sub> has been provided by Alfa Aesar and olivine by Minelco. Olivine has been calcined at 900°C for 10 h prior to use in the gasification reaction to enhance its reactivity for tar cracking. The conditions mentioned for olivine calcination have been determined as optimum by Devi et al. [14] in order to maximize tar cracking activity. The BET surface area has been measured by N<sub>2</sub> adsorption-desorption (Micromeritics ASAP 2010). Calcined olivine has a limited porosity, with a surface area of only 0.18 m<sup>2</sup> g<sup>-1</sup>. However,  $\gamma$ -Al<sub>2</sub>O<sub>3</sub> has a much higher porous development, with a surface area of 159 m<sup>2</sup> g<sup>-1</sup>.

### 2.4 Product analysis

The volatile stream leaving the gasification reactor has been analyzed online by means of a GC Agilent 6890 provided with a HP-PONA column and a flame ionization detector (FID). The sample has been injected into the GC by means of a line thermostated at 280°C, once the reactor outlet stream has been diluted with an inert gas. The purpose of this system is to avoid the condensation of tars in the transfer line. The tars collected in the condensation system have been identified in a gas chromatograph/mass spectrometer (GC/MS, Shimadzu UP-2010S provided with a HP-PONA column). The non-condensable gases have been injected into a micro-GC (Varian 4900).

### 2.5 Experimental procedure

Temperature and steam/biomass ratio are the operating parameters studied in the gasification of biomass and plastics in this reactor. Additionally, biomass gasification was also performed with different primary catalysts (in situ), and the influence of using the fountain confiner was evaluated. In all runs, water flow rate was 1.5 mL min<sup>-1</sup>, corresponding to a steam flow rate of 1.86 L min<sup>-1</sup>, which is approximately 1.5 times that corresponding to the minimum spouting velocity in order to ensure bed stability.

The effect of temperature has been studied at 800, 850, and 900°C by feeding a mass flow rate of 1.5 g min<sup>-1</sup> of biomass or HDPE and using a steam/feed ratio of 1.

The effect of the steam/feed ratio has been studied between 0 and 2 (in mass), and the temperature has been maintained at 900°C. For a ratio of 2, the biomass or



plastic feed rate was reduced to  $0.75 \text{ g min}^{-1}$  in order to maintain the same steam flow rate ( $1.5 \text{ g min}^{-1}$ ). The reactor contains 70 g of sand in the bed in all runs, and therefore the residence time of the products in the reactor and the hydrodynamic behavior are similar. In order to study the steam/feed ratio of 0, the steam was replaced with a  $\text{N}_2$  flow rate of  $2 \text{ L min}^{-1}$ .

In the experiments to assess the effect of the primary catalyst on product distribution, the bed contains 70 g of sand or olivine (with particle diameter in the 0.35–0.4 mm range). However, given that  $\gamma\text{-Al}_2\text{O}_3$  has a much lower density, the bed of this material contained 25 g with a particle size greater than that of sand, in the 0.4–0.8 mm range, in order to attain a similar hydrodynamic behavior in all cases. The experiments were carried out at  $900^\circ\text{C}$ , with a feed rate of  $1.5 \text{ g min}^{-1}$  of HDPE or sawdust and with a steam/feed ratio of 1.

In the experiments performed with the fountain-confined spouted bed, the biomass feed rate was  $0.75 \text{ g min}^{-1}$ , with a steam/biomass ratio of 2. The bed contained 100 g of olivine, and two particles sizes have been used, i.e., 90–150 and 250–355  $\mu\text{m}$ . These olivine particle size ranges are those corresponding to the optimum hydrodynamic performance of the reactor, as the minimum spouting velocity depends strongly on particle size [33]. Thus, the gas velocity in the runs with the coarse olivine fraction corresponds to approximately 1.5 times the minimum spouting velocity (so the reactor operated under conventional spouting regime), whereas in the experiments performed with the fine olivine, the gas velocity used is approximately four times higher than the minimum spouting velocity ( $4 u_{\text{ms}}$ ), and the fountain-enhanced regime was therefore attained.

Furthermore, operation was carried out in two regimes in the same reactor in order to ascertain the influence the confinement system (in the standard spouting regime) has on the biomass gasification process. Thus, experiments with and without the fountain confiner were carried out at  $850^\circ\text{C}$  and S/B of 2, using coarse olivine (250–355  $\mu\text{m}$ ), with gas velocity corresponding in both cases to approximately 1.5 times  $u_{\text{ms}}$  (conventional spouting regime). The results obtained with the confiner under conventional spouting regime were compared with those obtained with this device but operating in the enhance fountain spouting regime under the same conditions and replacing the coarse olivine with the fine one in the bed. Therefore, the role of the vigorous gas-catalyst contact in the fountain-enhanced regime was assessed.

All the runs were performed in continuous mode for 20 min in order to ensure a steady-state process. The char yield was determined by weighing the mass in the reactor, as well as those retained in the cyclone and in the sintered steel filter. The char yield is given by mass unit of the whole amount of solid fed into the reactor (approximately 30 g). All the runs have been repeated several times (at least three) under the same conditions in order to guarantee reproducible results.

### **3. Results and discussion**

#### **3.1 HDPE gasification**

In this work, steam gasification of HDPE has been studied in the conventional conical spouted bed pilot plant described in Section 2.2. The effect of temperature (in the  $800\text{--}900^\circ\text{C}$  range) and steam/plastic (S/P) ratio (between 0 and 2) on the gas yield, tar content, carbon conversion efficiency, and  $\text{H}_2$  production is shown in **Table 2**. The reaction indices have been defined as follows: (i) gas yield as the volumetric gas production (on a dry basis) per kg of biomass in the feed (on a wet

basis), (ii) tar yield expressed as the tar mass per syngas  $\text{m}^3$  (on a dry basis), (iii) carbon conversion efficiency as the ratio between the carbon units contained in the syngas and those contained in the biomass in the feed, and (iv)  $\text{H}_2$  production as the mass percentage of the  $\text{H}_2$  produced per biomass mass unit.

The gaseous fraction is composed of  $\text{H}_2$ ,  $\text{CO}$ , and  $\text{CO}_2$ , together with  $\text{C}_2$ – $\text{C}_5$  hydrocarbons (mainly  $\text{C}_3$ –). The tar is defined as the amount of organic compounds with a molecular weight and boiling point higher than that of benzene, a criterion that is commonly used by most authors [11, 14, 34]. The char is a carbonaceous product collected after the reaction in the reactor, sintered steel filter, and cyclone. The mass balance closure in all the experiments was above 95%.

As observed in **Table 2**, an increase in temperature leads to higher gas yields and lower tar and char yields, thus improving the efficiency of the whole process. The gas yield increases from  $2.5 \text{ m}^3 \text{ kg}^{-1}$  of HDPE at  $800^\circ\text{C}$  to  $3.4 \text{ m}^3 \text{ kg}^{-1}$  of HDPE at  $900^\circ\text{C}$ . Furthermore, the carbon conversion efficiency at  $800^\circ\text{C}$  is 86%, increases to 91% at  $850^\circ\text{C}$ , and then remains constant with further increases in temperature to  $900^\circ\text{C}$ .

Tar content decreases from  $29.5 \text{ g Nm}^{-3}$  at  $800^\circ\text{C}$  to  $16.7 \text{ g Nm}^{-3}$  at  $900^\circ\text{C}$  due to the enhancement of thermal cracking. Other authors have also observed a positive effect of temperature on the tar cracking in the gasification of waste plastics by using both steam [35] and air [36] as gasifying agents. In fact, according to certain authors, the destruction of tar aromatic hydrocarbons only occurs at temperatures above  $850^\circ\text{C}$  [13].

The influence of temperature on product yields has also been studied with different gasification technologies, and most of the authors agree that higher temperatures enhance syngas yield and decrease that of tar and char [37–39]. Higher char yields than those shown in **Table 2** have been reported in the literature [35, 40], which may be attributed to the characteristics of the gas-solid contact in the conical spouted bed reactor, which mitigate the limitations in the physical steps prior to gasification, which are as follows: (i) plastic melting, (ii) coating of sand particles, and (iii) pyrolysis.

**Table 2** also displays the reaction indices for different S/P values. As observed, as S/P ratio is increased from 1 to 2, the carbon conversion increases from 91.0 to 93.6%. Note that the performance is poor when operating with a S/P = 0 (pyrolysis), given that carbon conversion efficiency is as low as 68.6% due to the high tar and char yields. The lack of steam in the reactor at high temperatures promotes the formation of aromatic compounds, leading to a tar content as high as  $29.5 \text{ g Nm}^{-3}$ . The presence of steam in the reaction medium increases the gas yield and decreases that of tar. When operating only with  $\text{N}_2$  as a fluidizing agent, the tar concentration

Temperature ( $^\circ\text{C}$ )	S/P ratio	Tar content ( $\text{g Nm}^{-3}$ )	Carbon conversion (%)	Gas yield ( $\text{m}^3 \text{ kg}^{-1}$ )	$\text{H}_2$ production (wt%)	Char yield (wt%)
800	1	29.5	86.1	2.5	12.7	1.4
850	1	13.8	91.1	3.2	17.0	0.6
900	1	16.7	91.1	3.4	18.4	0.5
900	2	9.6	93.6	3.6	19.9	0.4
900	0	207.8	68.6	0.9	2.7	5.6

**Table 2.** Effect of gasification temperature and S/P ratio on the gas yield, tar content, carbon conversion efficiency, and  $\text{H}_2$  production.

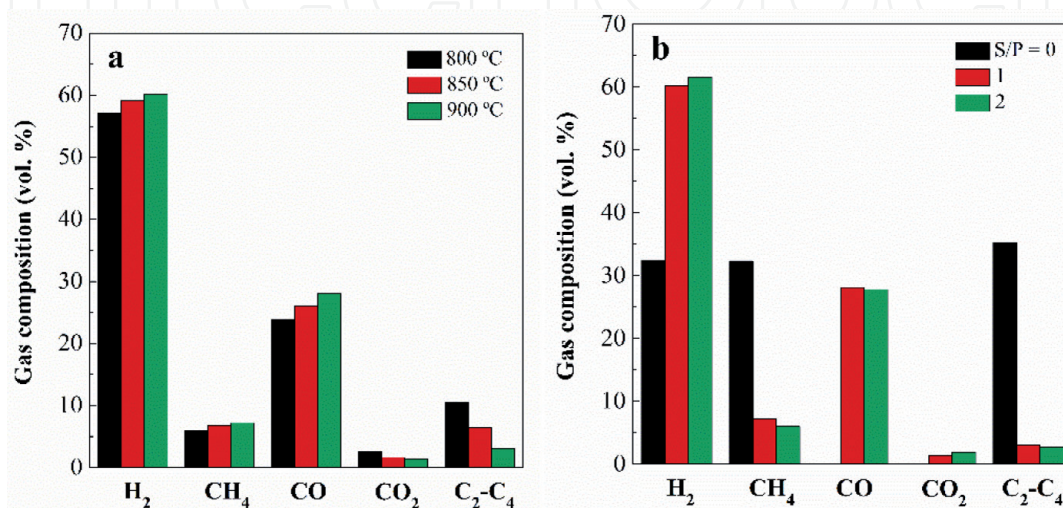
is  $207.8 \text{ g Nm}^{-3}$ , but this concentration is drastically reduced to  $16.7 \text{ g Nm}^{-3}$  and  $9.6 \text{ g Nm}^{-3}$  when operating with S/P ratios of 1 and 2, respectively. These results suggest that an increase in S/P ratio enhances the cracking of tar compounds, as reported by Herguido et al. [41] in the steam gasification of biomass.

The presence of steam in the reaction environment also improves  $\text{H}_2$  production, increasing significantly from 2.7 to 18.4 wt% when the S/P ratio is increased from 0 to 1. However, the increase in  $\text{H}_2$  production (19.9 wt%) is moderate when a S/P value of 2 is used. Similarly, gas yield increases slightly from  $3.4 \text{ m}^3 \text{ kg}^{-1}$  HDPE to  $3.6 \text{ m}^3 \text{ kg}^{-1}$  HDPE when the S/P ratio is raised from 1 to 2. The following aspects can explain these results: (i) promotion of hydrocarbon reforming reactions (Eq. (1)) as steam concentration is higher and (ii) low tar and char formation rate, although this effect is of lower significance. A similar trend has been reported in the literature, although some authors attain a saturating trend, i.e., a higher steam/tire ratio than the optimum one does not increase the gas yield [42, 43].

Moreover, **Figure 3** displays the composition of the gases formed at different temperatures (**Figure 3a**) and S/P ratios (**Figure 3b**). As observed in **Figure 3a**, an increase in temperature leads to an increase in the concentrations of  $\text{H}_2$ , CO, and  $\text{CH}_4$  in the gaseous stream, which are 60.3, 28.2, and 7.2% vol., respectively, at  $900^\circ\text{C}$ . Temperature has an opposite effect on  $\text{C}_2\text{--C}_5$  hydrocarbons (made up mainly of olefins, with ethylene being the major one), whereas that on  $\text{CO}_2$  was almost negligible (the concentration is almost steady).

The higher concentration of  $\text{H}_2$  and CO can be explained by the endothermic nature of steam and dry reforming reactions (Eqs. (1) and (4)), which are promoted at higher temperatures, whereas that of  $\text{CH}_4$  is due to the endothermicity of HDPE cracking reactions. On the contrary, the  $\text{C}_2\text{--C}_5$  hydrocarbons formed are probably reformed, and therefore their yield decreases as temperature is higher. It should be noted that the water-gas shift reaction (Eq. (6)) is exothermic, and therefore thermodynamic equilibrium shifts toward the formation of CO at high temperatures.

Regarding the gas composition (**Figure 3b**), an increase in S/P ratio from 1 to 2 does not lead to a significant change, but the composition of the gas when only pyrolysis is performed (S/P = 0) is very different. As observed, the presence of steam favors  $\text{H}_2$  and  $\text{CO}_2$  formation but reduces that of CO and  $\text{CH}_4$  because the higher concentration of steam in the reactor enhances both water-gas shift and methane reforming reactions. Other authors have observed a similar effect of S/P ratio on the gas composition in the gasification of different polymeric materials [42, 43].



**Figure 3.** Effect of gasification temperature (a) and S/P ratio (b) on the gaseous fraction composition.



## 3.2 Biomass gasification

### 3.2.1 Effect of temperature and S/B ratio

The same reaction indices in plastic gasification, i.e., the gas yield, tar content, carbon conversion efficiency, and H<sub>2</sub> production, have been analyzed in this section (**Table 3**). Temperature is one of the more influential variables in steam gasification, and its effect has been studied in the 800–900°C range for a steam/biomass ratio of 1. Moreover, the effect of steam/biomass ratio has been studied in the 0–2 range at 900°C. Note that no steam was fed into the reactor in the runs carried out with a S/B ratio of 0, but the sawdust contained a moisture content of approximately 10%, and water is formed during the thermal degradation of biomass. Therefore, some steam reforming will occur even in the runs without water in the feed (runs with S/B = 0).

As observed in **Table 3**, the temperature plays a crucial role in the efficiency of the gasification process. An increase in the gasification temperature reduces the tar content in the gaseous product from 364.4 g Nm<sup>-3</sup> at 800°C to 142.5 g Nm<sup>-3</sup> at 900°C. The gas yield also increases from 0.7 m<sup>3</sup> kg<sup>-1</sup> of biomass at 800°C to 1 m<sup>3</sup> kg<sup>-1</sup> of biomass at 900°C, whereas that of char decreases from 8.9% at 800°C to 4.5% at 900°C. In the same line, the carbon conversion efficiency in the process is considerably higher as temperature is increased, and char yield is therefore lower. In fact, char gasification kinetics is enhanced by temperature due to the highly endothermic nature of char steam gasification (Eq. (3)) and Boudouard (Eq. (4)) reactions. The increase in char conversion with temperature is related to the shift in equilibrium in both reactions [44]. However, this result depends on the char residence time in the reactor. Thus, char gasification reaction kinetics is slow, even above 800°C.

Although the content of tar is reduced to 142.5 g Nm<sup>-3</sup> operating at 900°C due to the positive effect of temperature on tar cracking and reforming reactions, this value is still high for syngas applications. It should be noted that no defluidization problems are observed in the steam gasification, which is due to the vigorous solid cyclic movement in the conical spouted bed. However, the conventional spouted bed regime leads to short residence times (below 0.5 s), which are beneficial to increase the yield of bio-oil in pyrolysis processes, but in gasification they are responsible for the limited tar cracking, whose concentration in the gaseous stream is rather high, as observed in **Table 3**.

Given that the tar yield is highly dependent on several parameters, such as residence time, temperature, and S/B ratio, the results showed in the literature vary greatly depending on the technology used, but all of them evidence a significant decrease in tar content in the gaseous product stream with temperature [45–47].

With respect to the experiments carried out with different S/B ratios (**Table 3**), an increase in this parameter improves the gasification performance by increasing the gas yield and carbon efficiency and lowering that of tar. For example, tar concentration has been reduced from 154 g Nm<sup>-3</sup> with a S/B = 0 to 142.5 g Nm<sup>-3</sup> with a S/B = 1, given that an increase in the S/B ratio promotes tar cracking and reforming reactions (Eq. (1)). However, a further increase in the S/B ratio from 1 to 2 only reduces slightly the tar content of the gaseous product. Likewise, the gas yield increases from S/B 0 to 1 (from 0.9 to 1 m<sup>3</sup> kg<sup>-1</sup> of biomass) but hardly changes as S/B is increased from 1 to 2.

The reduction in the tar and char content leads to an increase in the carbon conversion efficiency, attaining the maximum value of 70% with a S/B = 2. Although gasification efficiency is improved in terms of biomass conversion, the energy efficiency of the process is lower when high S/B ratios are used, given that more water

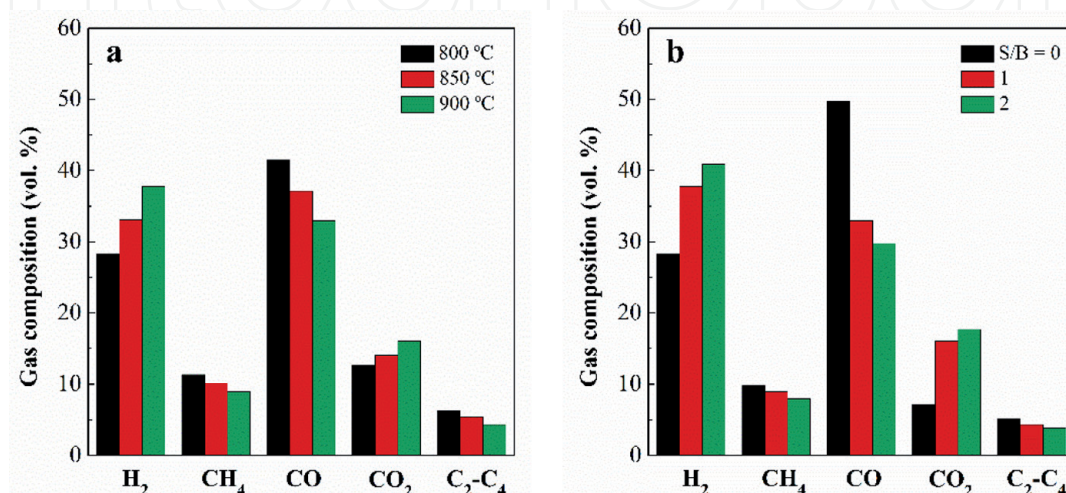
Temperature (°C)	S/P ratio	Tar content (g Nm <sup>-3</sup> )	Carbon conversion (%)	Gas yield (m <sup>3</sup> kg <sup>-1</sup> )	H <sub>2</sub> production (wt%)	Char yield (wt%)
800	1	364.2	50.4	0.7	1.9	8.9
850	1	243.1	59.1	0.8	2.5	6.3
900	1	142.5	69.8	1.0	3.2	4.5
900	2	142.0	70.0	1.0	3.6	3.6
900	0	154.0	50.4	0.9	2.3	10.7

**Table 3.** Effect of gasification temperature and S/B ratio on product fraction yields, carbon conversion efficiency, and tar concentration, at 900°C.

need to be vaporized and the unreacted steam needs to be recovered after being condensed. Kaushal and Tyagi [48] suggest optimum S/B ratios between 0.6 and 0.85, which guarantee the thermal efficiency of the process and, at the same time, the presence of enough steam in the gasifier to promote steam reforming reactions.

The composition of the gases (on a dry basis) formed at different temperatures and different S/B ratios is displayed in **Figure 4**. As observed in **Figure 4a**, an increase in temperature enhances H<sub>2</sub> formation due to the endothermic nature of the reactions involved (Eqs. (1)–(5)). Moreover, the inorganic species of the biomass retained in the char have a positive effect on the water-gas shift reaction (Eq. (6)) at higher temperatures [46]. Accordingly, H<sub>2</sub> concentration increases from 28% at 800°C to 38% at 900°C, whereas that of CO decreases from 41.5 to 32.5% in the same range of temperature. Besides, concentration of methane and the other gaseous hydrocarbons (C<sub>2</sub> to C<sub>4</sub>) decreases as temperature is raised due to the enhancement of hydrocarbon reforming reactions. As in HDPE gasification, the effect of temperature on CO<sub>2</sub> is not of significance, as its concentration increases slightly between 800 and 900°C.

**Figure 4b** shows the composition of the gaseous stream for different S/B ratios. Given that the WGS reaction and methane and hydrocarbon reforming reactions (Eqs. (1) and (2)) are promoted at high S/B ratios, the formation of H<sub>2</sub> and CO<sub>2</sub> is enhanced, whereas that of CO and hydrocarbons is hindered. It is to note that this effect is more remarkable when the S/B ratio is increased from 0 to 1.



**Figure 4.** Gas composition (on a dry basis) for the steam gasification at different temperatures (a) and S/B ratios (b).



### 3.2.2 Effect of primary catalyst

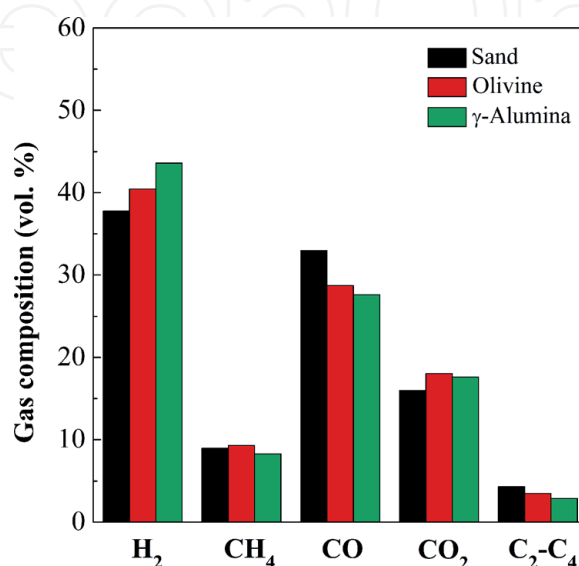
The experiments with different bed materials have been carried out at a temperature of 900°C and a S/B ratio of 1. **Table 4** shows the effect of the primary catalysts used (olivine and  $\gamma$ -alumina) on reaction indices (gas yield, tar content, H<sub>2</sub> production, and carbon conversion) and compares the results with those obtained using inert sand as bed material. As observed, both olivine and  $\gamma$ -alumina cause a great decrease in tar content (30.1 and 22.4 g Nm<sup>-3</sup>, respectively) compared to the runs carried out with inert sand (142.5 g Nm<sup>-3</sup>). Accordingly, both catalysts improve the gasification performance, with tar reduction being slightly higher for  $\gamma$ -alumina (84%) than that for olivine (79%). Moreover, the carbon conversion efficiency has a drastic increase when a primary catalyst is used, attaining a value of 86.8% for olivine and 87.6% for  $\gamma$ -alumina. It is noteworthy that H<sub>2</sub> production peaks at 4.5 wt% when the  $\gamma$ -alumina is used.

As mentioned above, tar formation leads to operational problems in the gasification and subsequent units for syngas processing; thus, the use of a catalyst, such as olivine and  $\gamma$ -alumina, improves process efficiency, especially the latter, which significantly reduces tar content. Nevertheless, olivine is cheaper and more available because it is a natural material [49]. Other papers in the literature also report considerable improvements in gasification efficiency by using primary catalysts [45, 50].

The effect primary catalysts have on gas composition is displayed in **Figure 5**. As observed,  $\gamma$ -alumina has a greater influence on gas composition than olivine. The presence of catalysts leads to an increase in H<sub>2</sub> and CO<sub>2</sub> concentrations and a

	Sand	Olivine	$\gamma$ -Alumina
Tar content (g Nm <sup>-3</sup> )	142.5	30.2	22.4
Carbon conversion (%)	69.7	86.8	87.6
Gas yield (m <sup>3</sup> kg <sup>-1</sup> )	1.0	1.1	1.2
H <sub>2</sub> production (wt%)	3.2	3.7	4.5
Char yield (g Nm <sup>-3</sup> )	4.5	4.3	4.3

**Table 4.**  
 Effect of the primary catalysts on reaction indices.



**Figure 5.**  
 Effect of primary catalysts on the composition of the gaseous fraction.

reduction in that of CO due to the promotion of the water-gas shift reaction (Eq. (6)). In addition, the higher concentration of H<sub>2</sub> by the presence of this type of catalyst is also related to the enhancement of tar cracking and reforming reactions (Eq. (1)). Moreover,  $\gamma$ -alumina also seems to promote methane and light hydrocarbon reforming (Eq. (3)), which can be deduced from their lower concentration in the presence of this catalyst.

### 3.2.3 Effect of fountain confinement on biomass gasification

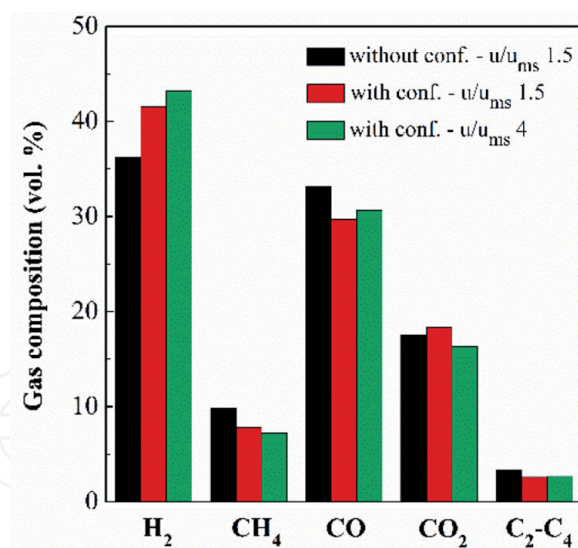
Runs have been carried out with a S/B ratio of 2 and at a temperature of 850°C with different spouting regimes and gas flow patterns developed in conical spouted beds, such as (i) standard spouting regime without fountain confiner, (ii) standard spouting regime with fountain confiner, and (iii) enhanced fountain regime with fountain confiner. **Table 5** compares the gas yield, tar content, carbon conversion efficiency, char yield, and H<sub>2</sub> production results obtained for the three configurations mentioned.

As observed in **Table 5**, the incorporation of the fountain confiner leads to a decrease in tar content in the syngas from 49.2 g Nm<sup>-3</sup> without fountain confiner to 34.6 g Nm<sup>-3</sup> when this device is inserted. The volatiles in the conventional spouted bed gasifier leave quickly from the reaction zone through the outlet located in the gasifier upper section. Thus, the short residence time of the volatiles limits the contact of tars and other gaseous products with the catalyst, which hinders cracking and reforming reactions and therefore lowers conversion efficiency. On the contrary, the fountain confiner prevents the premature leaving of the gases at an initial stage in the biomass gasification and causes a downward gas flow inside the confiner, which favors the contact between the volatile stream and the catalyst. Furthermore, the confined fountain and the use of draft tubes lead to a highly stable hydrodynamic regime, which allows operating with finer materials (lower particle sizes of olivine) and higher fountain heights [24].

In order to analyze the influence on the gasification performance by changing the gas-catalyst contact in the reactor, especially in the fountain region, runs with the fountain confiner were performed under similar residence times (same reactor geometry and gas flow rate) as in conventional conical spouted beds. As observed in **Table 5**, the promotion of steam reforming of tars and gaseous hydrocarbons using the confinement system improved the gas yield and H<sub>2</sub> production from 1.1 to 1.2 m<sup>3</sup> kg<sup>-1</sup> and from 3.5 to 4.6 wt%, respectively. In the same line, the carbon conversion efficiency also increased when the confinement system was used, given that a value of 83.6% was obtained instead of 81.5% without this system. It should be remarked that these values are slightly higher than those reported by other authors in fluidized bed reactors under similar conditions [51, 52].

	Without confiner	With confiner (standard spouting)	With confiner (enhanced fountain)
Tar content (g Nm <sup>-3</sup> )	49.2	34.6	20.6
Carbon conversion (%)	81.5	83.6	86.1
Gas yield (m <sup>3</sup> kg <sup>-1</sup> )	1.1	1.2	1.3
H <sub>2</sub> production (wt%)	3.5	4.6	5.0
Char yield (g Nm <sup>-3</sup> )	6.5	6.2	6.0

**Table 5.**  
Influence of the confinement system and spouting regime on the reaction indices.



**Figure 6.**  
*Influence of the confinement system and spouting regime on gas composition.*

**Table 5** also shows that the results are greatly improved under fountain-enhanced regime by decreasing olivine particle size and increasing the fountain height. In fact, the tar content in the gas is reduced from  $34.6 \text{ g Nm}^{-3}$  under conventional spouting regime up to  $20.6 \text{ g Nm}^{-3}$  under enhanced fountain regime. This improvement is associated with the better gas-catalyst contact and heat transfer rates in the fountain region due to the higher fountain height. Furthermore, the smaller particle size of olivine increased the catalyst surface area available for cracking and reforming reactions [14]. Moreover, gas composition with and without confiner (under conventional and fountain-enhanced regime) is shown in **Figure 6**.

As observed in **Figure 6**, H<sub>2</sub> concentration increases from 36 to 42% with and without the fountain confiner, whereas that of CO decreases. The effect on CO<sub>2</sub> is not so remarkable, but its concentration is slightly higher when the fountain confiner is introduced. Furthermore, the concentration of methane and the other gaseous hydrocarbons decreased due to the higher extent of steam reforming reactions involving methane (Eq. (2)) and tar (Eq. (1)), as well as of water-gas shift (Eq. (6)) reactions when the fountain confiner was used. This improvement is related to the increase in the gas residence time and the better contact of the gas with the catalyst attained when the fountain confiner is used. It is noteworthy that effect of the fountain-enhanced regime on the gas composition is rather limited. The most significant change is that regarding H<sub>2</sub> concentration, whose value increases to 43.2%.

#### 4. Conclusions

The conical spouted bed reactor is an interesting technology for the continuous steam gasification of biomass and waste plastics due to the high heat transfer rates for a highly endothermic process (as is gasification) as well as to the absence of defluidization problems. An increase in gasification temperature improves process efficiency in terms of conversion to gases, with the maximum carbon conversion being of 70 and 91.1% at 900°C for biomass and HDPE, respectively. Furthermore, steam/feed ratio has a positive effect on the composition of the gas by increasing the H<sub>2</sub> concentration from 32 to 61% in the HDPE gasification and from 28 to 42% in that of biomass when steam/feed ratio is increased from 0 to 2. In fact, higher steam concentrations in the reaction environment enhance both tar cracking and char

gasification and so increase carbon conversion efficiency. Nevertheless, the concentration of the tars attained is still high for its direct application. The use of primary catalysts, such as olivine and  $\gamma$ -alumina, has shown an excellent performance for tar elimination as their content is being reduced by up to 30.1 and 22.4 g Nm<sup>-3</sup> with olivine and  $\gamma$ -alumina, respectively.

The incorporation of a fountain confiner in the CSBR allows modifying bed hydrodynamics, i.e., increase the residence time of the volatiles and improve their contact with the catalyst in order to promote gasification performance and favor tar cracking. Hence, H<sub>2</sub> productions and carbon conversion efficiencies increase when the fountain confiner is introduced from 3.5 to 4.6 wt% and from 81.5 to 83.6%, respectively. Moreover, the H<sub>2</sub> concentration increases from 36 to 42%, whereas that of CO decreases from 34 to 29% with and without the fountain confiner. This device allows operating under enhanced fountain regime by reducing olivine particle size, which leads to a better contact between olivine and the gases, and therefore tar content is further reduced, and the carbon conversion efficiency increases up to 86.1%.

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