Experimental study and model validation of waste gasification in an up-draft fixed-bed gasifier

N. Verdone, P. De Filippis, M. Scarsella & B. de Caprariis
Sapienza University of Rome, Italy

Abstract

Gasification has been identified as a key technology to enhance the environmental tolerability of low quality fuels such as waste and biomass. In this work, the performances of a laboratory scale gasification process fed with waste are reported. Among the several technical choices, the up-draft fixed-bed gasifier was selected as an interesting solution for heat generation in small-scale applications, due to the characteristics of simple geometry and low cost. The experimental setup is composed of an up-draft gasifier followed by a reactor used as filter to remove the particulate and as second thermal and catalytic stage to convert the produced tar to lighter species. A literature model has been adapted to the case under study to analyse the influence of operative parameters such as oxidant flow rate (equivalent and air/steam ratio values) and gasification temperature of the process. The original literature model considers the species gas evolution along the axial coordinate only and does not include time dependency. To make the model time dependent, the consumption time of the gasification fuel bed estimated from experiments was introduced. Furthermore the modelling of the oxidation zone was introduced, adding the char combustion equation. Since the model concerns only the gasification, the up-draft process was split into two consecutive steps to allow direct comparison between experimental and simulated data: first the drying and pyrolysis processes and then the fixed bed gasification. During the pyrolysis, a second stage reactor for the tar reforming was included, allowing a tar reduction of about 85%. The model was successfully validated with experimental data and then was used to predict the operative parameters that determine the optimal syngas composition. The best syngas composition (35% CO and 10% H₂) was obtained with an equivalent ratio of 0.6 and a bed temperature of 1100 K.

Keywords: auto fluff, disposal, gasification, energy recovery.
1 Introduction

Automotive shredder residue (ASR) or auto fluff is generated by technologies that recover non-ferrous metals and plastics from end-of-life vehicles, appliances and other light scrap. The main flow of ASRs is generated by the shredding of end-of-life vehicles (ELVs). The ASR is principally composed of textiles (25%), plastic (23%), sponge (17%), as well as sand and solids (16%) [1]. The ASR represents 20% of the ELVs that is not recycled and is disposed in landfill in the majority of the EU countries [2]. Sometimes, the presence of heavy metals and chloride from metals and PVC respectively makes ASR a hazardous waste causing increasing costs for its disposal.

The incineration is another possibility for ASRs disposal, but the high content of inert leads to a low grade heating value making difficult the combustion of the ASR alone. Usually the co-combustion with municipal solid waste is preferred.

The quantity of ASR is forecast to rise in the coming years because of the increasing amount of polymers replacing metals in new vehicles. ASR valorisation treatments are necessary according to the limit of 13 MJ/kg stated for waste disposal by the European Directive 1999/31/EC and according to the Directive 2000/53/EC that claims for ELVs at least 85% recycling rate and 95% recovery rate by the year 2015, thus allowing the disposal in landfill for maximum 5% ELV total weight [3]. Thus energy recovery from ASR will become one of the most attractive possibilities to dispose auto fluff.

In this work, it is proposed to recover energy from auto fluff by gasification. In the gasification process, a combustible gas from a solid fuel with low grade heating value is produced. The up-draft fixed-bed gasifier was selected as an interesting solution for heat generation due to the characteristics of simple geometry and low cost.

A one-dimensional model was implemented to simulate the behaviour of the gasification stage. The composition of the produced gas was estimated and compared with experimental data showing a good agreement. The model was used to optimize the operative conditions of the gasifier in order to obtain the best syngas composition.

2 Experimental set-up

2.1 Experimental apparatus

A laboratory-scale gasification plant was designed and constructed. Its main components are schematically represented in fig. 1. The plant is composed of an updraft gasifier followed by a catalytic reforming reactor, where the tar cracking occurs and the dragged particulate gasification reactions are completed. The catalyst used in the reforming reactor is made of aluminium oxide. The two reactors are cylindrical steel tubes of 4 cm diameter \((D)\) and 60 cm length \((L)\). The high value of the \(D/L\) reactor ratio makes it difficult to have an autothermal process, therefore the two reactors are heated by external tubular ovens. A grate, placed at 10 cm from the bottom of the gasifier, is used as support for the solid
combustible and as air distributor. The fuel can be injected continuously through a cochlea whereas the ash removal is discontinuous. The two gasification agents, steam and air, are premixed, preheated and then injected into the reactor in counter-current with respect to the solid flow.

Figure 1: Laboratory scale plant.

Temperature profiles along the gasifier axis are measured by four thermocouples placed in a ceramic protective tube. The CO, CO$_2$ and CH$_4$ concentrations are monitored continuously. The H$_2$ concentration is measured with a gas chromatograph every 15 min. The plant is equipped with two gas outflows: at the exit of the gasifier and at the exit of the reforming reactor. A condenser for the collection of the vapours produced during the pyrolysis is placed after the reforming reactor.

2.2 Materials

ASR is a heterogeneous material composed of a mixture of plastics, rubber, resins and textiles and by a small amount of glass fragment and unrecovered metals. Because of the heterogeneity of this material, samples were first accurately ground and mixed. The immediate and elemental analyses were performed and the calorific value was calculated experimentally with a Mahler bomb calorimeter. Average values of these properties are reported in table 1.

3 Model

A literature model [4] for a down-draft gasifier was adapted to our case. The model is one-dimensional, the bed and gas properties variations were considered negligible and the gases were assumed to behave ideally.
Five reactions were taken into consideration [5]:

1) \( C + O_2 \rightarrow CO_2 \)
2) \( C + CO_2 \rightarrow 2CO \)
3) \( C + H_2O \rightarrow CO + H_2 \)
4) \( C + 2H_2 \rightarrow CH_4 \)
5) \( CH_4 + H_2O \rightarrow CO + 3H_2 \)

<table>
<thead>
<tr>
<th>Table 1: ASRs properties.</th>
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</thead>
<tbody>
<tr>
<td>Immediate Analysis</td>
</tr>
<tr>
<td>------------------------</td>
</tr>
<tr>
<td>Humidity (% mass)</td>
</tr>
<tr>
<td>Volatile matter (% mass dry basis)</td>
</tr>
<tr>
<td>Fixed carbon (% mass dry basis)</td>
</tr>
<tr>
<td>Ash (% mass dry basis)</td>
</tr>
<tr>
<td>Upper calorific value (MJ/kg)</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Elemental Analysis</th>
<th>(% mass dry basis)</th>
</tr>
</thead>
<tbody>
<tr>
<td>C</td>
<td>20</td>
</tr>
<tr>
<td>H</td>
<td>6.1</td>
</tr>
<tr>
<td>O</td>
<td>16</td>
</tr>
<tr>
<td>N</td>
<td>0.5</td>
</tr>
<tr>
<td>S</td>
<td>0.3</td>
</tr>
<tr>
<td>Cl</td>
<td>&lt; 0.1</td>
</tr>
</tbody>
</table>

The first reaction is the char combustion reaction, which takes place at the bottom of the gasifier where the air is injected. The gasification zone follows the combustion one. The gasification reactions are endothermic. The heat necessary to activate these reactions is in part supplied by the exothermic combustion and in part by the furnace.

The reaction rates were considered to follow an Arrhenius temperature dependence and to be proportional to the difference between the actual and the equilibrium reactant concentration:

\[
r_1 = C_{RF,G}A_1 n \exp\left(-\frac{E_1}{RT}\right) \left( P_{O_2} - \frac{P_{CO_2}^2}{K_1} \right) \tag{1}
\]

\[
r_2 = C_{RF,G}A_2 n \exp\left(-\frac{E_2}{RT}\right) \left( P_{CO_2} - \frac{P_{CO}^2}{K_2} \right) \tag{2}
\]

\[
r_3 = C_{RF,G}A_3 n \exp\left(-\frac{E_3}{RT}\right) \left( P_{H_2O} - \frac{P_{CO}P_{H_2}}{K_3} \right) \tag{3}
\]

\[
r_4 = C_{RF,G}A_4 n \exp\left(-\frac{E_4}{RT}\right) \left( P_{H_2}^2 - \frac{P_{CH_4}^2}{K_4} \right) \tag{4}
\]

\[
r_5 = C_{RF,G}A_5 n \exp\left(-\frac{E_5}{RT}\right) \left( P_{CH_4}P_{H_2O} - \frac{P_{CO}P_{H_2}^3}{K_5} \right) \tag{5}
\]

where \( A \) and \( E \) [J.mol\(^{-1}\)] are the reaction Arrhenius parameters, \( n \) [mol m\(^{-3}\)] the molar concentration of gaseous species, \( C_{RF} \) the char reactivity factor, which was
estimated from experimental data as described below and $K_j$ the equilibrium constants for each reaction that were calculated from the Gibbs energy of reaction:

$$\ln K_j = -\frac{\Delta G_j}{RT}$$

The constants $C_{RF,C}$ and $C_{RF,G}$ are the char reactivity factors for the combustion and the gasification reactions, respectively. These parameters represent the reactivity of different types of char. They affect the production rate of the species and the temperature profile in the reactor. These parameters were considered different for the combustion and the gasification reactions and were estimated by fitting the temperature experimental curve as explained in the next section.

The chemical species considered in this work are CO, CO$_2$, CH$_4$, H$_2$, H$_2$O, N$_2$, and O$_2$. For each species $i$ the mass balance equation is:

$$\frac{d(n_i \cdot v)}{dz} = R_i$$

where $z$ is the bed axial coordinate, $v$ [m.s$^{-1}$] the superficial gas velocity and $R_i$ [mol.m$^{-3}$.s$^{-1}$] is the net rate of formation and is reported in table 2.

### Table 2: Net rate of production of the different gaseous species

<table>
<thead>
<tr>
<th>Species</th>
<th>$R_i$ (mol/m$^3$.s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>N$_2$</td>
<td>0</td>
</tr>
<tr>
<td>O$_2$</td>
<td>$-r_1$</td>
</tr>
<tr>
<td>CO$_2$</td>
<td>$r_1 - r_2$</td>
</tr>
<tr>
<td>CO</td>
<td>$2 \cdot r_2 + r_3 - r_5$</td>
</tr>
<tr>
<td>CH$_4$</td>
<td>$r_4 - r_5$</td>
</tr>
<tr>
<td>H$_2$O</td>
<td>$-r_3 - r_5$</td>
</tr>
<tr>
<td>H$_2$</td>
<td>$r_3 - 2 \cdot r_4 + 3 \cdot r_5$</td>
</tr>
</tbody>
</table>

The variation of the temperature ($T$ [K]) on the axial coordinate was carried out using an energy balance:

$$\frac{dT}{dz} = \frac{1}{v \cdot \sum_i n_i c_i} \left( -\sum_j r_j \Delta H_j - v \frac{dp}{dz} - P \frac{dv}{dz} - \sum_i R_i \cdot c_i \cdot T \right)$$

(6)

where $-\sum_j r_j \Delta H_j$ is the energy released by the chemical reaction $j$, $c_i$ [J.mol$^{-1}$K$^{-1}$] the molar heat capacity of the species $i$ and $P$ [Pa] the total pressure.

Two further equations are needed to complete the reactor modelling: an equation for the pressure gradient along the bed and one for the variation of the superficial gas velocity. The pressure gradient is a function of the superficial gas velocity and it was derived from the Ergun equation [6]:

$$-\frac{dp}{dz} = \frac{150 \mu (1 - \varepsilon)^2}{D_p^2 \varepsilon^3} v + \frac{175 \rho (1 - \varepsilon)}{D_p \varepsilon^3} \cdot v^2$$

(7)

where $\varepsilon$ is the void fraction of the bed, $D_p$ [m] the particle diameter, $\mu$ [kg.m$^{-1}$s$^{-1}$] the fluid viscosity and $\rho$ [kg.m$^{-3}$] the fluid mass density.
The expression for the velocity gradient was obtained by differentiating the ideal gas law:

\[
\frac{dv}{dz} = \frac{1}{\sum_i n_i c_i + n R} \left( \frac{\sum_i n_i c_i \sum_i R_i}{n} - \frac{\sum_j r_j \Delta H_j}{T} - \frac{dP}{dz} \left( \frac{v}{T} + \frac{v \sum_i n_i c_i}{p} \right) - \sum_i R_i c_i \right)
\]  

(8)

A system of ten coupled first order differential equations were obtained including ten variables: concentrations of the seven considered species, temperature, pressure and velocity.

To make the model time dependent, the variation of the bed length as a function of time was calculated experimentally, as explained in the next session.

### 3.1 Char reactivity factor and bed length consumption rate determination

To determine the bed length consumption rate and the CRF factor, experiments on a smaller tubular quartz reactor heated by an electrical resistance were performed. The same operative conditions, scaled to the size of the smaller reactor, were reproduced. Four thermocouples were disposed at 0, 0.5, 1.2 and 2.6 cm along the reactor axial coordinate. The bed consumption rate (\(\alpha\)) was calculated as:

\[
\mathcal{R}(t) = -\alpha t + H_0
\]

where \(\mathcal{R}(t)\) and \(H_0 = 5\) cm are the bed length as a function of time and the bed length initial value, respectively.

A curve of the temperature profile along the reactor axial coordinate is reported in fig. 2 for three different times. The time needed for the complete bed consumption was calculated considering that the end of gasification occurs when the temperature is almost constant in the reactor. The curve at 250 s represents the beginning of the process. The combustion process occurs where the temperature is higher. Then the temperature decreases since the endothermic gasification reactions take place. At \(t = 2,000\) s, the temperature is almost constant along the bed height and the process is considered finished.

A fitting to find the correct value of the two CRF factors was made on the temperature profile. In fig. 2, the comparison between experimental and simulated data is reported showing a good agreement. The CRF values were calculated to be 100 and 1500 for the combustion and gasification reactions, respectively.

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**Figure 2:** Temperature profile along the gasifier axis at three different residence time, experimental and simulated data.
4 Results and discussion

In an up-draft fixed-bed gasifier, the combustible and the oxidant flow counter-currently, the fuel is introduced at the top of the gasifier and the oxidant at the bottom. In fig. 3, a generic representation an up-draft gasifier with the typical temperature profile is reported.

![Figure 3: Up-draft gasifier and temperature profile [7].](image)

When the fuel is injected, it is dried and then the pyrolysis process occurs. The char forms a fixed bed in the reactor where the gasification and the combustion reactions take place. The combustion zone is the first zone from the bottom of the reactor and the exothermic reactions make the process autothermal. The raw gas exits at the top of the gasifier and is laden with pyrolysis products such as tar, oil and light hydrocarbons. Thus to complete the gasification reactions and to allow the tar cracking, a reforming unit is generally placed after the gasifier.

Two types of experiments were performed. First a complete gasification, using the reactor as an updraft gasifier was carried out. In fig. 4, the concentration profiles of CH\textsubscript{4}, CO and CO\textsubscript{2} of a complete updraft gasification process are reported. It can be noticed that the pyrolysis and the gasification processes occur in two separated steps: the pyrolysis process takes place in the first 1000 s, i.e. during methane production. The gasification occurs until after the CO concentration is higher than that of CO\textsubscript{2}. The CO concentration becomes smaller than the CO\textsubscript{2} one as the bed height decreases and the CO\textsubscript{2}, produced by the combustion, does not have enough time to react with the residue char.

![Figure 4: Concentration profile of a complete up-draft gasifier.](image)
Since experimental data on the gasification process alone were needed, the process was split into two parts: the pyrolysis carried out at 850°C in an inert atmosphere and the gasification in a fixed bed.

The results of the pyrolysis are reported in table 3. The volatile concentrations were calculated before and after the reforming reactor. The amount of tar collected before the reforming reactor was calculated to be 30% of the volatile matter and the hydrogen concentration is 9%. The concentration of the pyrolysis gas species as a function of time after the reforming unit is reported in fig. 5.

Table 3: Pyrolysis gas and tar concentration.

<table>
<thead>
<tr>
<th></th>
<th>CO</th>
<th>CO₂</th>
<th>CH₄</th>
<th>H₂</th>
<th>Tar</th>
</tr>
</thead>
<tbody>
<tr>
<td>Before</td>
<td>21%</td>
<td>15%</td>
<td>11.4%</td>
<td>9%</td>
<td>30%</td>
</tr>
<tr>
<td>After</td>
<td>25%</td>
<td>15%</td>
<td>16%</td>
<td>19%</td>
<td>5%</td>
</tr>
</tbody>
</table>

Figure 5: Concentration profile of CH₄, CO and CO₂ during the olive residue pyrolysis after the reforming reactor.

After the reforming unit, a great reduction of the tar amount that is converted in lighter species can be noticed. The reforming unit converts most of the tar enriching the gas in hydrogen and methane. Thus the reforming reactor is fundamental to reduce the tar quantity that can create problems in the following zones of the plant.

The model was validated using experimental data. In fig. 6, the comparison between simulated and experimental data is reported, showing a good agreement.

The model was used to predict the gas composition in order to optimize the operative conditions that gave the best syngas composition. In table 4, the simulated gas compositions obtained by varying the ER, the steam air ratio and the temperature are reported.

Increasing the amount of steam and the temperature and decreasing the ER ratio, the quality of the obtained syngas improves. However, using a low equivalent ratio can be a problem since the gasification process becomes too slow. As expected, increasing the steam flow, the amount of hydrogen rises. This effect could not be convenient because the gasification time increases, the reactions including the water vapour are endothermic, and the steam production is an expensive operation that can reduce consistently the plant efficiency. The best gas composition is CO 35%, CO₂ 2.25% and H₂ 11.5% that is a typical gas composition for air gasification.
Figure 6: Experimental CO (blue line) and CO2 (red line) profile compared with CO (green line) and CO2 (violet line) simulated profile obtained for a ER = 0.6, T = 1100 K and 30% of steam.

Table 4: Simulated results of the syngas composition expressed in % in volume.

<table>
<thead>
<tr>
<th>ER</th>
<th>T (K)</th>
<th>O2 steam/O2 air</th>
<th>%CO</th>
<th>%CO2</th>
<th>%H2</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.5</td>
<td>1000</td>
<td>10%</td>
<td>25.73</td>
<td>4.75</td>
<td>8.00</td>
</tr>
<tr>
<td>0.6</td>
<td>1000</td>
<td>10%</td>
<td>23.56</td>
<td>5.49</td>
<td>6.50</td>
</tr>
<tr>
<td>0.7</td>
<td>1000</td>
<td>10%</td>
<td>20.19</td>
<td>6.50</td>
<td>4.30</td>
</tr>
<tr>
<td>0.6</td>
<td>1100</td>
<td>10%</td>
<td>30.45</td>
<td>2.30</td>
<td>8.50</td>
</tr>
<tr>
<td>0.7</td>
<td>1100</td>
<td>10%</td>
<td>26.54</td>
<td>4.35</td>
<td>5.70</td>
</tr>
<tr>
<td>0.6</td>
<td>1100</td>
<td>20%</td>
<td>33.15</td>
<td>2.60</td>
<td>9.75</td>
</tr>
<tr>
<td>0.7</td>
<td>1100</td>
<td>20%</td>
<td>29.30</td>
<td>4.10</td>
<td>6.80</td>
</tr>
<tr>
<td>0.6</td>
<td>1100</td>
<td>30%</td>
<td>35.00</td>
<td>2.25</td>
<td>11.50</td>
</tr>
</tbody>
</table>

5 Conclusions

The gasification represents promising technology to recover energy from waste when the calorific value is low. The pyrolysis process produced a good quality gas that can be used as fuel and the reforming unit reduces considerably the tar amount that can cause problem in the following part of the plant. The syngas contains a low quantity of pollutant and can be sent to a purifying unit and then used directly as a combustible or as a raw material for other processes like the hydrogen production or the Fischer Tropsch. The composition of the obtained syngas is in line with typical syngas composition obtained by means of air gasification.

References


