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# CO-GASIFICATION OF BIOMASS AND COAL IN AN 8 MW DUAL FLUIDIZED BED STEAM GASIFIER

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## ABSTRACT

Gasification of biomass is an attractive technology for combined heat and power production. Co-gasification of biomass and coal was tested in an 8 MW dual fluidized bed steam gasifier with coal ratios up to 22% on an energy basis. Hydrogen levels in the producer gas increased with the addition of coal as well as ammonia, hydrogen sulfide and tars. Addition of coal to the system stabilized the process and improved gas quality.

## INTRODUCTION

Coal as a substitute for oil is nothing new. During the oil crises in the 1970s and up until the middle of the 1980s, coal was already being used as a substitute for oil (1). But in these days, not only was an oil substitute required but also a way to minimize the carbon dioxide emissions, which many countries agreed to in the Kyoto protocol. Since carbon dioxide emissions from biomass are perceived as being neutral (1), and since coal is a fuel with a high availability (especially in politically stable countries) and is less expensive than oil, the gasification of mixtures of these two oil alternatives is a natural consequence. The availability of biomass fluctuates mostly with the season (2), hence the idea of gasifying miscellaneous mixtures in one plant looks economically advantageous. Generally, co-firing is the use of different fuels at the same time for combustion or gasification. For example, biomass is co-fired in existing coal plants with fuels that cannot be burned alone because of the low energy content (such as sewage sludge), and it could be burned together with natural gas to give a good performance. In industrial coal-fired power plants co-firing is often used to add a green fingerprint without any loss in efficiency and with only minor changes to plant settings. Therefore, only low percentages of the other fuels are usually used. Co-combustion of biomass with coal is a matter of intensive research for different applications and several comprehensive studies exist on this topic (3-11). Different types of reactors can be used such as fixed bed, fluidized bed, dual fluidized bed and entrained flow reactors. Co-firing can be accomplished via three different modifications, which are direct, indirect and parallel co-firing. The first two are favorable since for indirect co-firing a separate gasification unit is required. Direct co-firing uses blends of both fuels, and for parallel co-firing the fuels are fed into separate boilers to produce steam. Different kinds of fuels have been investigated, such as agricultural residues together with coal (12). Moreover, the use of energy plants such as *Cynara* (13) and sewage sludge (14) are options for the co-firing of biomass feedstocks together with coal. Recently, the co-gasification of biomass and coal has attracted more interest due to the environmental benefits such as reduced sulfur and nitrogen emissions when adding biomass to the fuel for systems designed for coal. Moreover, CO<sub>2</sub> emissions can also be reduced. Several reports about co-gasification are available (15-19). However, only a small amount of literature is available for tests with different ratios of biomass and coal (2,20-21).

Within this work the suitability of coal for the dual fluidized bed gasification process was tested in a commercially operated 8 MW plant in Güssing, Austria. In order to guarantee a positive test run at the plant, test runs at the 100 kW scale were previously undertaken (22-23). These tests showed that blend ratios of 0 to 100% were possible. Thus, the goal of the tests described in this paper was to verify the findings from the pilot scale to the large scale. Due to the limitations (see the Results section) of existing plants, the maximum possible coal ratio was 22% in terms of energy. However, the general aim was to demonstrate the fuel flexibility of the dual fluidized bed process.

## EXPERIMENTAL

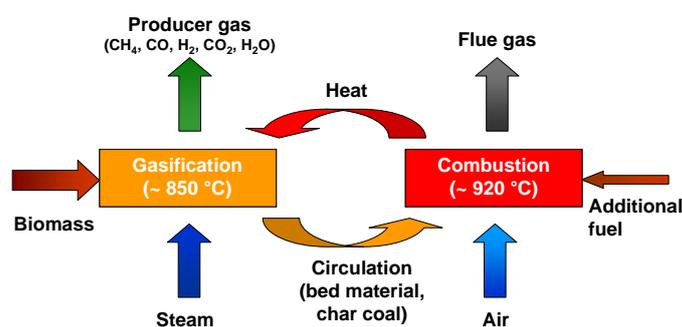


Figure 1: Principles behind the dual fluidized bed gasifier

Figure 1 shows the principles behind the dual fluidized bed steam gasification process and Figure 3 shows a schematic of the process. The biomass enters a bubbling fluidized bed gasifier where drying, thermal degasification, and partially heterogeneous char gasification take place at bed temperatures of about 850-900 °C. The residual biomass char leaves the gasifier together with the bed material through an inclined, steam fluidized chute, towards the combustion reactor. The combustion reactor is used to heat up the bed material and is designed as a highly expanded fluidized bed (riser). Air is used as a fluidization agent in the riser. After particle separation from the flue gas in a cyclone, the hot bed material flows back to the gasifier via a loop seal. Both connections, between the loop seal and the chute, are fluidized with steam, which effectively prevents gas leakage between the gasification and combustion zones and also allows high solid throughput. The temperature difference between the combustion and the gasification reactors is determined by the energy needed for gasification as well as the circulation rate of the bed material. The system is inherently auto-stabilizing since a decrease in the gasification temperature leads to higher amounts of residual char, which results in more fuel for the combustion reactor. This, in turn, transports more energy into the gasification zone and thereby stabilizes the temperature. In practical operations, the gasification temperature can be influenced by the addition of fuel (e.g. recycled producer gas, sawdust, etc.) into the combustion reactor. The pressure in both gasification and combustion reactors is close to atmospheric conditions.

The process yields two separate gas streams at high temperatures: a high quality producer gas and a conventional flue gas. The producer gas is generally characterized by a relatively low content of condensable higher hydrocarbons (2-10 g/m<sup>3</sup> of so called tars, heavier than toluene), low N<sub>2</sub> (< 1 vol%db), and a high H<sub>2</sub> content of 35-40 vol%db. For practical use, Olivine, a natural mineral, has been proven to be a suitable bed material with enough resistance to attrition and moderate tar cracking activity.

## A combined heat and power plant in Güssing, Austria

The demonstration plant in Güssing was developed based on the results gathered from a 100 kW<sub>Fuel input</sub> pilot plant at Vienna University of Technology. The fuel power of the demonstration plant is 8 MW<sub>Fuel input</sub>, the electrical output is 2 MW and the thermal output 4.5 MW. A simplified flow sheet is shown in Figure 2. More than 42,000 hours of combined heat and power (CHP) operation has been achieved since it was commissioned in 2002. Wood chips from the forestry are used as fuel. The producer gas is cooled and cleaned by a two-stage cleaning system. A water cooled heat exchanger reduces the temperature from 850-900 °C to about 160-180 °C. The first stage of the cleaning system is a fabric filter to separate the particles and some of the tar from the product gas. These particles are returned to the combustion zone of the gasifier. In the second stage, the gas is liberated from the tar by a scrubber. At the same time, the product gas is cooled down to about 40 °C, which is necessary for the gas engine. The spent scrubber liquid saturated with tar and condensate is vaporized and fed into the combustion zone of the gasifier for thermal disposal. If the gas engine is not in operation the entire amount of producer gas can be burned in the boiler to produce heat. The sensible heat of the engine's flue gas is used to produce district heat, and the flue gas from the combustion zone is used for preheating air and superheating steam, as well for delivering heat to the district heating grid.

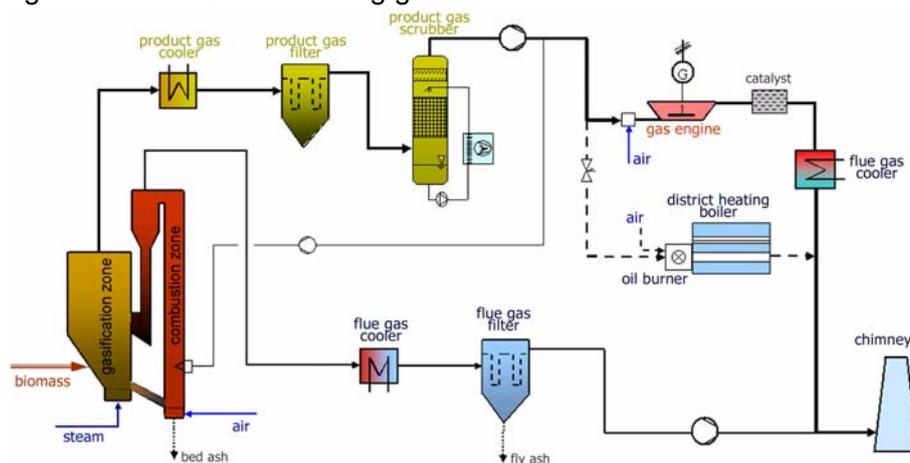


Figure 2: Simplified flow sheet of the combined heat and power plant in Güssing, Austria

## Analysis

The components CH<sub>4</sub>, H<sub>2</sub>, CO, and CO<sub>2</sub>, as well as O<sub>2</sub>, were measured by a Rosemount NGA 2000. The components N<sub>2</sub>, C<sub>2</sub>H<sub>4</sub>, C<sub>2</sub>H<sub>6</sub>, and C<sub>3</sub>H<sub>8</sub> were measured via an online gas chromatograph. Minor contaminants (H<sub>2</sub>S, NH<sub>3</sub>) and tar were measured between the producer gas filter and the scrubber. The tars were sampled isokinetically with washing bottles with toluene as the absorption liquid and gravimetric as well as GC/MS tars were determined. For the ammonia measurement, gas was sampled in a similar way to the tar measurements, using washing bottles. The solvent used in this procedure was diluted sulfuric acid at a temperature of about 2 °C. Hydrogen sulfide was again sampled first using washing bottles filled with an aqueous potassium hydroxide solution at a temperature of about 2 °C. Subsequently, the H<sub>2</sub>S values were determined by potentiometry. The flue gas was measured with a Rosemount NGA 2000 (CO, CO<sub>2</sub>, O<sub>2</sub>, and NO).

Sulfur dioxide ( $\text{SO}_2$ ) was measured by potentiometry after sampling with washing bottles with an aqueous potassium hydroxide solution as the absorption liquid at a temperature of about  $2\text{ }^\circ\text{C}$ . Details about the measurement methods can be found elsewhere (22).

## RESULTS

Throughout the whole test campaign all of the main parameters were kept constant; where this was not possible it will be mentioned in the following discussion. The bed pressure in the gasification section was adjusted to 120 mbar, the steam-to-fuel ratio to  $0.67\text{ kg}_{\text{steam}}/\text{kg}_{\text{dry fuel}}$  (which corresponded to  $1.35\text{ kg}_{\text{steam}}/\text{kg}_{\text{carbon}}$ ) and the gasification temperature was  $870\text{ }^\circ\text{C}$  by default. A schematic of the gasification reactor and a regime map of the gas/solid contact are given in Figure 3 and Figure 4, respectively. A comprehensive description of the gas/solid contact in a fluidized bed reactor was previously given (24-25).

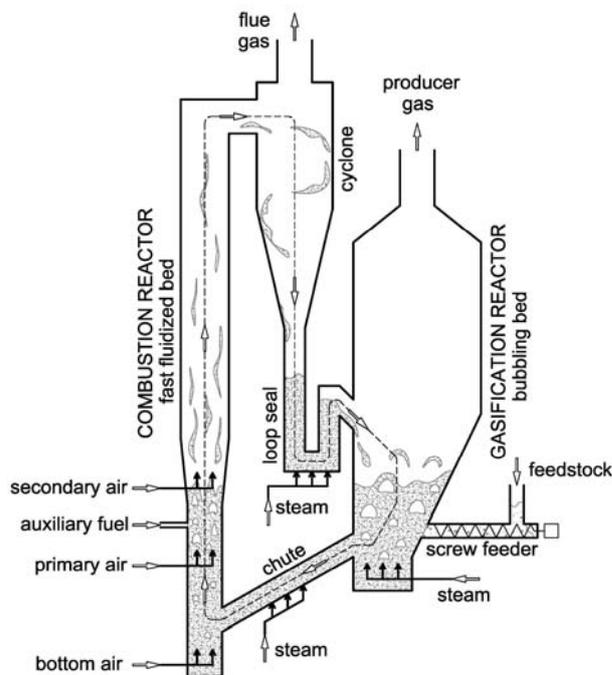


Figure 3: Process schematic of the dual fluidized bed gasifier in Güssing, Austria

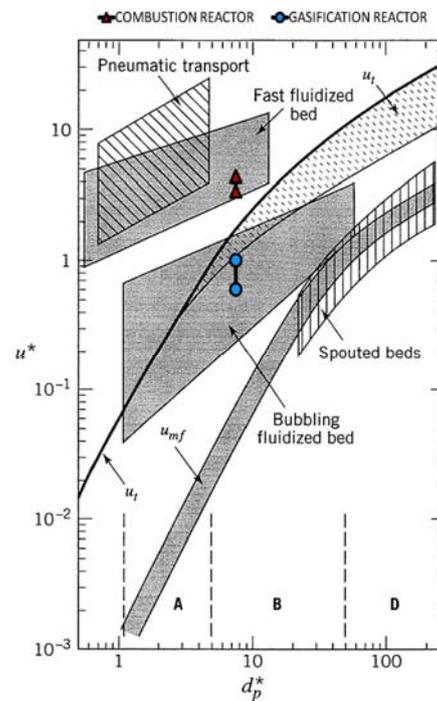


Figure 4: Regime map of the dual fluidized bed gasifier in Güssing, Austria (25)

It can be seen that the regime in the gasification section of the reactor is a bubbling bed, whereas the combustion section lies in the fast fluidization section. The operational area is determined by the fact that producer gas is produced in the bed over the height and in the combustion section air is introduced at three different levels. The superficial gas velocity in the riser after the secondary inlet ranges from 10 to 13 m/s, whereas superficial gas velocities of 1.8 to 3 m/s occur in the gasification zone. The minimum fluidization velocity  $U_{mf}$  for both reactors (product gas at  $850\text{ }^\circ\text{C}$  and flue gas at  $920\text{ }^\circ\text{C}$ ) varied from 0.11 to 0.13 m/s, and the terminal velocity  $U_t$  was in the range of 4.6 to 5.3 m/s.

Untreated wood chips from the forestry were used (mainly hard wood) as a standard fuel for the plant. Polish hard coal was added at ratios of 12, 18, and 22% on an energy basis. The proximate and ultimate analyses of the fuels are listed in Table 1. Each operation point was applied for at least one day in order to obtain reliable figures. It should be mentioned that due to an accumulation of coal in the system the gasification temperature slightly increased at the operation point with 22% of coal. However, at the pilot scale it was possible to operate the gasifier even with 100% of coal, whereas the load had to be reduced due to the low reaction rate of coal at these temperatures (23).

|                          | Polish hard coal |             | Wood chips |             |
|--------------------------|------------------|-------------|------------|-------------|
|                          | Dry basis        | As received | Dry basis  | As received |
| Water content            | -                | 6.11        | -          | 27.7        |
| Ash content              | 2.89             | 2.76        | 1.0        | 0.94        |
| C                        | 82.17            | 78.43       | 48.82      | 46.06       |
| H [mass %]               | 4.57             | 4.36        | 5.87       | 5.54        |
| N                        | 1.66             | 1.58        | 0.15       | 0.14        |
| O                        | 8.08             | 12.27       | 44.16      | 47.32       |
| S                        | 0.14             | 0.13        | 0.005      | 0.003       |
| Volatile matter [mass %] | 34.68            | 33.1        | 84.02      | 79.27       |
| Fixed carbon [mass%]     | 65.32            | 62.35       | 15.98      | 15.08       |
| LHV [MJ/kg]              | 31.6             | 30.1        | 18.2       | 17.0        |

Table 1: Proximate and ultimate analyses of the fuels

Figure 5 shows the main producer gas components and Figure 6 shows the higher hydrocarbons as well as the lower heating value LHV vs. the coal ratio. Hydrogen, carbon monoxide and methane slightly increased whereas carbon dioxide decreased. Surprisingly, the opposite trend was found for carbon monoxide in the pilot plant (fluidization 2010). Ethane ( $C_2H_6$ ) and propane ( $C_3H_8$ ), showed no significant trend, whereas ethene ( $C_2H_4$ ) seemed to decrease slightly, which corresponds to the findings at the pilot scale ( $100 \text{ kW}_{\text{Fuel input}}$ ) (22). Moreover, the amount of char transported with the bed material to the combustion zone increased due to the lower reaction rate of coal in comparison to the biomass. Hence, more thermal energy bound in char was available in the combustion part of the facility and less additional fuel needed to be burned in the combustor. This is in fact the limitation for the coal ratio at the existing plant since the gas burners are cooled by the gas and the volume flow cannot be reduced under a certain limit. Generally, the dual fluidized bed gasifier can also handle 100% coal as fuel (22-23), whereas due to the lower reaction rate of coal in comparison to biomass the load had to be removed. Another option would be to increase the residence time of the char fraction in the gasification section to increase the conversion rate by changing the geometry and/or the bed material circulation rate. In summary, the process was stabilized since pressure fluctuations due to the devolatilization of the biomass were reduced. As described above, two different tar measurements were taken via GC/MS as well as gravimetrically. Three samples were taken daily between the producer gas filter and the scrubber (see Figure 2). With an increase in the coal ratio the gravimetric tar and the GC/MS tar increased significantly, as shown in Figure 7. The same trends, although less distinctive, were found at the pilot scale. Figure 8 shows the nitrogen and sulfur mass flows into the gasifier via the fuel versus the applied coal ratio. The mass flows of nitrogen released as  $NH_3$  and sulfur released as  $H_2S$  with the producer gas are displayed. Linear trends were measured for both

impurities. Nearly all sulfur ended up in the producer gas as  $H_2S$ , whereas only about 50% of the nitrogen inlet flow was transferred to ammonia. This correlation was previously found for the dual fluidized bed steam gasification technology (26).

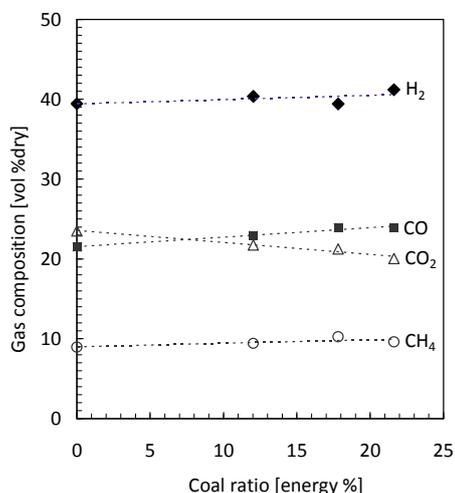


Figure 5: Main producer gas components vs. coal ratio

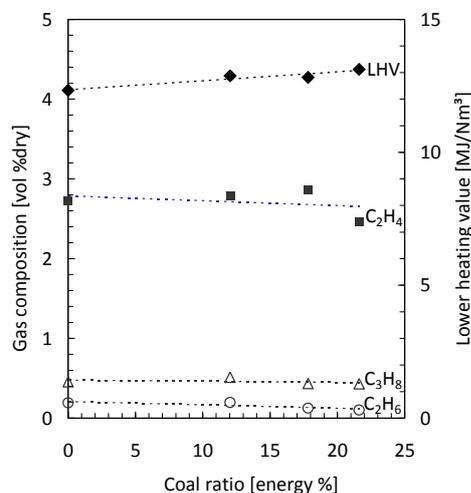


Figure 6: Higher hydrocarbons and lower heating value vs. coal ratio

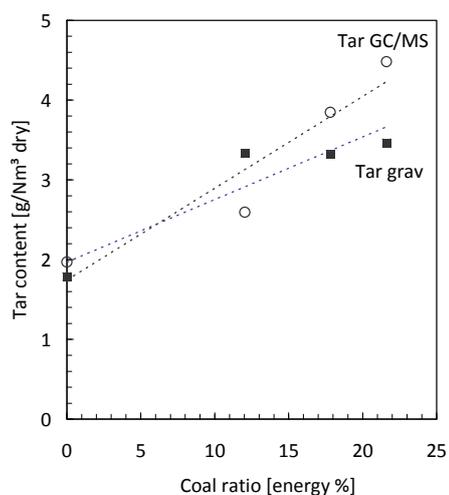


Figure 7: Gravimetric and GC/MS tars in the producer gas vs. coal ratio

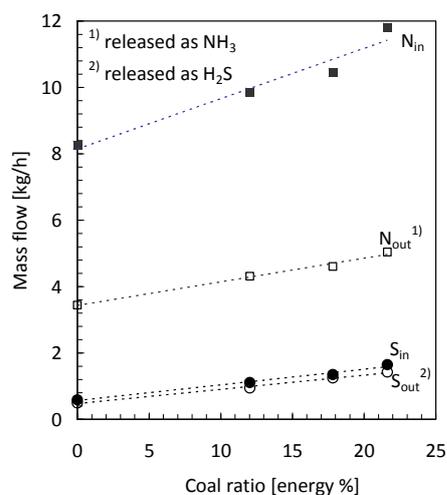


Figure 8: Nitrogen and sulfur mass flows in/out vs. coal ratio

## CONCLUSIONS

The experiments showed that coal can be added to the biomass as fuel for the dual fluidized bed steam gasification process. The gas composition shifted to higher hydrogen and carbon monoxide contents, which increased the lower heating value of the gas. Thus, the  $H_2/CO$  ratio can be adjusted to the needs of the applied utilization route (e.g. synthetic natural gas synthesis). The addition of up to 22% coal on an energy basis could be applied without major operational problems. During co-gasification the process was stabilized due to the lower reaction rate of coal as well

as to the reduced level of devolatilization. Tar levels in the producer gas slightly increased. Ammonia and hydrogen sulfide linearly increased with the addition of coal due to the higher nitrogen and sulfur contents in coal in comparison to biomass. Generally, the dual fluidized bed system offers excellent fuel flexibility for use in advanced power cycles as well as in future liquid/gaseous fuel production systems.

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## NOTATION

$U_{mf}$  minimum fluidization velocity, m/s  
 $U_t$  terminal velocity, m/s

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