# Steam gasification of biomass in dual fluidized bed gasifiers: A review

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ABSTRACT

Indirect or allothermal gasification of biomass in dual fluidized bed (DFB) gasifiers such as the Güssing gasifier or the biomass heatpipe reformer becomes particularly attractive for the conversion of biomass into hydrogen or any second generation fuel such as substitute natural gas (SNG), methanol or Fischer-Tropsch diesel fuel. Interconnected and indirectly heated DFB gasifiers produce syngas with $H_2/CO$ ratios of 2–3 and hydrogen concentrations even above 50 vol%(dry basis). Fluidized bed particles, the operating pressure, solids circulation rate and heat transfer coefficients determine the layout of these gasifiers. This article summarizes the state of the art with respect to layout and dimensioning of DFB gasifiers and reviews the impact of the steam equivalence ratio, fuel and bed material properties, char conversion, and combustion efficiency on cold gas efficiency and syngas quality of DFB gasifiers.

## 1. Introduction

Steam gasification in Dual Fluidized Bed (DFB) gasifiers is the most promising technology for the production of hydrogen and syngas from solid fuels in medium and small-scale applications. Large-scale plants for coal gasification will apply high pressure high temperature oxygen-blown entrained flow gasification due to the favorable gas quality or oxygen-blown fixed bed gasifiers if olefins and low-weight aromatic compounds are desired as by-products [1]. However, conversion of biomass requires smaller units. There are potentially situations where a local demand of hydrogen or methanol might foster a biomass-based production route. Low market prices for base chemicals such as hydrogen [2] or methanol [3] make small-scale biomass applications dependent on political incentives. Such incentives have been put in place for combined heat and power production in some European countries such as Germany and Austria, where several DFB gasifier plants have been built and operated since 2001.

Steam gasification of biomass and wastes provides significantly higher gas qualities than air-blown gasification in particular in the small and medium-scale range. Higher heating values and hydrogen contents up to 50% or even more make the technology particularly interesting for the conversion of biomass into second generation fuels such as Fischer-Tropsch diesel, methanol, dimethyl ether (DME) or substitute natural gas (SNG) [4]. Synthesis of these energy carriers requires nitrogen free syngas. State-of-the-art processes for the conversion of bituminous coal or lignite in large-scale units commonly use oxygen-blown gasification technologies. Since pure oxygen supply requires a considerable electric energy input and the efficiency of air separation units is better at large scale, oxygen-blown gasifiers are typically not applied in the power range for decentralized conversion of biomass. Hence, steam gasification is an alternative for processes using biomass as feedstock.

The key challenge with steam gasification is the necessity to provide the heat-of-reaction for the fuel drying, heating, and the endothermic gasification reactions at highest temperatures above 700–800 °C. Air- or oxygen blown gasifiers – so-called autothermal gasifiers – provide this heat-of-reaction by means of exothermal oxidation reactions, directly inside the gasifier. This *direct* or *autothermal gasification* either dilutes the produced gas with nitrogen in case of air-blown gasification or requires pure oxygen and costly air separation units. *Indirect* or *allothermal gasification* in oxygen free atmosphere requires heating of the gasification reactor [5].

Allothermal gasifiers usually use steam as gasification agent. An external heat source is needed to maintain the required operating temperature in the gasification reactor. Typical operating temperatures of allothermal biomass gasifiers range from 670 °C to 900 °C and are determined by chemical reaction constraints regarding carbon conversion, tar reforming, possible inherent capture of sulphur species and

## Nomenclature

| | |
|---|---|
| a, $a_{min}$ | (stoichiometric) air demand ($kg_{air}$ /$kg_{fuel}$) |
| A | cross section (of fluidized bed) ($m^2$) |
| Ar | Archimedes number |
| b | fuel-to-bed material ratio ($kg_{bed\ material}$/$kg_{fuel}$) |
| $\bar{c}_p$ | average specific heat capacity in relevant temperature range ($kJ\,kg^{-1}\,K^{-1}$) |
| $d_i$ | mean particle diameter of particle size class I (m) |
| $d_p$ | particle diameter (m) |
| $d_{sv}$ | Sauter mean particle diameter (m) |
| g | gravitational acceleration ($m\,s^{-2}$) |
| $G_s$ | solidx flux in riser ($kg\,m^{-2}\,s^{-1}$) |
| H | height of reactor, height of fluidized bed (m) |
| $H_l$ | lower heating value ($kJ\,kg^{-1}$) |
| $H_i$ | molar enthalpy ($kJ\,kmol^{-1}$) |
| $\Delta H_r$ | heat-of-reaction ($kJ\,kmol^{-1}$) |
| $\dot{m}$ | mass flow ($kg\,s^{-1}$) |
| $\dot{n}$ | molar flow ($kmol\,s^{-1}$) |
| $\widetilde{M}_i$ | molar mass ($kg\,kmol^{-1}$) |
| m, n | molar fraction of hydrogen and oxygen in the fuel ($kmol\,kmol^{-1}$) |
| $P_{th}$ | thermal fuel power based on lower heating value (MW) |
| $\Delta p$ | pressure drop along fluidized bed height (Pa) |
| $\dot{Q}$ | energy flow, heat demand, heat transfer rate (kW) |
| $\dot{q}$ | specific heat demand ($kW\,kW^{-1}$) |
| $\dot{Q}_{trans,circ}$ | heat transferred with circulating bed material (kW) |
| $\Delta\dot{Q}_{\Delta h,r}$ | heat-of-reaction for the endothermic gasification reactions (kW) |
| $\Delta\dot{Q}_{\Delta h,v}$ | heat-of-evaporation for the evaporation of the fuels moisture (kW) |
| $\Delta\dot{q}_{\Delta h,r}$ | specific heat-of-reaction for the endothermic gasification reactions (kW/kW) |
| $\Delta\dot{q}_{\Delta h,v}$ | specific heat-of-evaporation for the evaporation of the fuels moisture (kW/kW) |
| Re | Reynolds number |
| s, $s_{min}$ | (stoichiometric) steam demand ($kg_{steam}$ /$kg_{fuel}$) |
| T | thermodynamic temperature (K) |
| t | relative temperature (°C) |
| u | superficial gas velocity (m/s) |
| w | moisture content of fuel ($kg_{water}$ /$kg_{fuel}$) |
| $w_{H2O}$ | water content of fuel ($kg_{H2O}$ /$kg_{fuel}$) |
| $w_{ash}$ | ash content of fuel ($kg_{ash}$ /$kg_{fuel}$) |
| $x_{D,min}$ | minimum steam demand ($kg_{steam}$ /$kg_{fuel}$) |
| $x_i$ | mass fraction of particle size class i |
| X | fluidization factor |

*Greek letter*

| | |
|---|---|
| ε | voidage / porosity of fluidized bed ($m^3$/ $m^3$) |
| $\eta_{cg}$ | cold gas efficiency (kW/kW) |
| $\eta_{comb}$ | combustor efficiency (kW/kW) |
| $\eta_G$ | dynamic gas viscosity (Pa s) |
| λ | air equivalence ratio ($kg_{air}$ / $kg_{air,min}$) |
| $\rho_G$ | gas density ($kg\,m^{-3}$) |
| $\rho_P$ | particle hydrodynamic density ($kg\,m^{-3}$) |
| σ | steam equivalence ratio ($kg_{steam}$ / $kg_{steam,min}$) |
| $\nu_I$ | stoichiometric coefficient (kmol/kmol) |
| $\varphi_{char}$ | char conversion rate (kmol/kmol) |
| $\varphi_s$ | particle sphericity |
| Ψ | void fraction ($m^3$/$m^3$) |

*Subscripts*

| | |
|---|---|
| a | air |
| bm | bed material |
| char | char |
| circ | referring to bed material in circulation |
| comb | combustion chamber |
| f | fuel |
| fg | flue gas |
| HP | heat pipe |
| In | input |
| mf | at minimum fluidization conditions |
| p | referring to particle (diameter) |
| s | steam |
| sens | sensible heat |
| sg | syngas |

$CO_2$, as well as by agglomeration and fouling through ash melting depending on the fuel composition.

Three solutions are known for a sufficiently effective heating of such gasifiers:

1. Discontinuous intermittent operation of a single fluidized bed
2. Circulation of particles between two interconnected fluidized beds
3. Indirectly heated fluidized beds

All these technologies usually apply fluidized beds. Several design options for such steam gasifiers are summarized by Levenspiel [6]. Most of the currently proposed concepts combine fluidized bed combustion with fluidized bed gasification – comprising so-called dual fluidized bed (DFB) gasifiers. DFB gasifier systems are most appropriate to enable highest heat fluxes into relatively small gasification reactors: They allow effective heating by means of the circulation of hot solids or efficient indirect heating with integrated heat exchangers.

The term "DFB" usually applies to interconnected fluidized bed gasification systems. However, since characteristics and performances discussed in this review equally apply to interconnected and indirectly heated gasifier systems that comprise both, fluidized bed combustion and a fluidized bed gasification reactor, the term will be used for indirectly heated gasifier systems consisting of two hydraulically separated fluidized beds, too.

The main advantage of dual fluidized bed gasifier systems is the separation – at least with respect to the gas phases – of the gasification reaction and the combustion reaction required to supply the heat of reaction to the gasification reaction. This concept enables indirect gasification without dilution of the produced syngas with the nitrogen introduced with the combustion air.

Syngas heating values and hydrogen content of these allothermal gasifiers are comparable to autothermic gasification operated with steam-oxygen mixtures. Thus, allothermal steam gasifiers generate a syngas comparable to steam/oxygen blown gasifiers without the need for an air separation unit.

An alternative to the application of fluidized beds for indirect gasification of biomass with steam could be steam-blown plasma gasification [7]. In this case, the heat-of-reaction comes completely from electricity, which might be particularly interesting for any upcoming power-to-gas or power-to-fuel technology.

The general aim of the present article is to give an overview on the current state of knowledge about allothermal fluidized bed steam gasifiers for biomass working according to the DFB principle. Based on a detailed description of historic developments it will in particular discuss layout and design considerations for decentralized and small- to medium-scale applications. The specific objectives constitute in the