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Methodology for Minimising the Utility Consumption of a 2G Ethanol Process

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The production of ethanol from lignocellulosic biomass has gained increased interest in recent years, notably in the context of valorising agricultural by-products and providing fuels from renewable sources In order to increase their competitiveness, the energy demand of such processes needs to be minimised. This issue procures two benefits : (1) reduce utility consumption and (2) increase cogeneration possibility.
In the present article we investigate this problem for a study process : ethanol production from sugarcane bagasse by enzymatic hydrolysis and glucose fermentation. We therefore apply a rigorous optimisation methodology in which we control certain design parameters in order to maximize the net production of utility. As a result, we obtain a design for our process which (1) eliminates the need for an external hot utility, (2) minimizes the need for the cold utility and (3) maximises the cogeneration possibility.
As a conclusion, the proposed methodology provides a strong tool for minimising the utility consumption for a 2G ethanol plant. Considering its key components, it can further be applied in the context of a multi-objective problem.

1. State of the art analysis

Recent related research work has mainly dealt with the integrated 1st and 2nd generation ethanol production. (Dias et al. 2014) provides an overview of this subject and discusses some of its technical, economic and environmental aspects. (Mogensen et al. 2012) provides a techno-economic analysis and comparison of a handful of production scenarios. Mainly, the extent of heat and process integration as well as power cogeneration were evaluated. (Furlan et al. 2012) goes however a step further by coupling process simulation with a global optimisation algorithm. The goal of this work was to determine the optimal fraction of bagasse to be diverted to second generation ethanol production with regards to maximization of revenue. Moreover, (Ensinas et al. 2013) make use of a similar tool, but with a bi-objective optimisation: maximising electricity production versus maximising ethanol production. Furthermore, this work incorporates heat integration into the optimisation problem. Both these works stress the importance of coupling a process simulator to a global optimisation algorithm and conclude that the inclusion of 2G ethanol production greatly increases the heat demands of the process. This present article expands on the work of (Ensinas et al. 2013), for a stand-alone bagasse to ethanol plant, with the objective of minimising the process's utility consumption for a fixed ethanol production rate. This choice was triggered by two motives: (1) tackle the design problems related with the proposed route, (2) highlight the extent of the proposed methodology.

2. Description of the study process with base configuration

The process is simulated using Aspen Plus®. Its layout is inspired from the work of (Ensinas et al. 2013) and is highlighted in Figure 1. In this figure, we find the main process steps as well as the main input, output and intermediary streams. The main co-products : xylose and solid cake are valorised to provide heat for the process. Auxiliary fuel is burnt in case of an excess heat demand. On the contrary, in case of a

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low heat demand, heat is converted to steam and electricity in a Rankine cycle. Finally, waste heat is dissipated in the cold utility. We find in Table 1 a brief description of the various steps along with the chosen technology and certain design parameters. At this level, only design parameters directly affecting energy consumption are included. The base configuration consists of an initial choice for the values of the operating parameters. These values are based on the works of Bessa et al. (2012) and later Ensinas et al. (2013). As can be seen in Table 1, the values of certain process inputs are unknown at the simulation level. The calculation of their values is the subject of the next step : heat exchange network design.

Table 1: Description of main process steps, with chosen technologies and main design parameters

Step	Description	Chosen Technology	Main Design parameters
Bagasse pretreatment	Breakdown of cell structure	SO_2 catalyzed steam explosion.	Pretreatment temperature : 190 °C
Hydrolysis	Cellulose converted to glucose	Enzymatic digestion	Solids Loading : 2 %
Juice concentration	Concentrate the diluted glucose stream	Multi-effect evaporation	Fraction of discarded water : 95.4 % Fraction of water leaving each effect : $\frac{1}{6}$ 1st evaporator Temperature : 120 °C Adjacent ΔT: 10 °C Number of evaporators : 6
Fermentation	Convert glucose to ethanol	Yeast fermentation	Amount of purged CO_2: 99.5 %
Ethanol concentration	Obtain azeotropic water/ethanol mixture	Double-effect distillation	Operating pressures : 1 atm Fraction of water leaving 1st column : 96 % N stages (2 columns): 32 & 30 Feed stages : 8 & 28 Ethanol recovery : 99 %
Ethanol dehydration	Obtain ethanol at 99.3 % purity	Azeotropic distillation	Amount of third product added : 80 % of produced ethanol Product recoveries : 99 % N stages (2 columns) : 36 & 6 Feed stages : 22 & 3
Handling of xylose stream	Energetic valorization	Biodigestion & gas turbine	Burner pressure : 10 atm Burner temperature : 1,330 °C Turbine temperature : 1,130 °C
Handling of cake stream	Energetic valorization	Cake burner	Solid moisture content : 66 % Burner temperature : 527 °C
Auxiliary burner	External heat source	Natural gas burner (without gas turbine)	Burner temperature : 1,330 °C $\dot{m}_{auxiliary\ fuel}$: unknown
Cold utility	External heat sink	Cold water	Input/Output T levels : 25 °C $\dot{m}_{cold\ utility}$: unknown
Heat Recovery	Recovery of heat in cogeneration system	Electricity cogeneration by the use of a Rankine cycle	Degrees of superheating: 200 °C Number of draw-offs : 5 $P_{draw\ offs}$: 16; 3.5; 2.05; 0.17 & 0.12 atm $\dot{m}_{draw\ off,i}$:: unknown

3. Heat exchange network (HEN) design

The heat exchange network design problem intervenes once the process has simulated and converged. It is a mono objective mixed integer linear optimisation problem (MILP), which seeks to determine a heat exchange design which optimises a given objective function, whilst respecting the heat balance.
The chosen objective function is the net production of utility, U_{Prod}, expressed in Eq(1)

$$U_{PROD} = W + E_{cold\_utility} + E_{auxiliary\_fuel} \text{ (MW)} \quad (1)$$

W is the net power produced by the system. $E_{cold\_utility}$ is the heat lost to the cold utility, always negative. $E_{auxiliary\_fuel}$ is the thermal content of the required auxiliary fuel, based on its lower heating value (LHV), also always negative. As a result of this step, we obtain values for the missing flow rate parameters, subsequently values for the power production and utility consumption and finally a value for U_{Prod}. This step also enables the representation of exchanges by the use of composite curves.

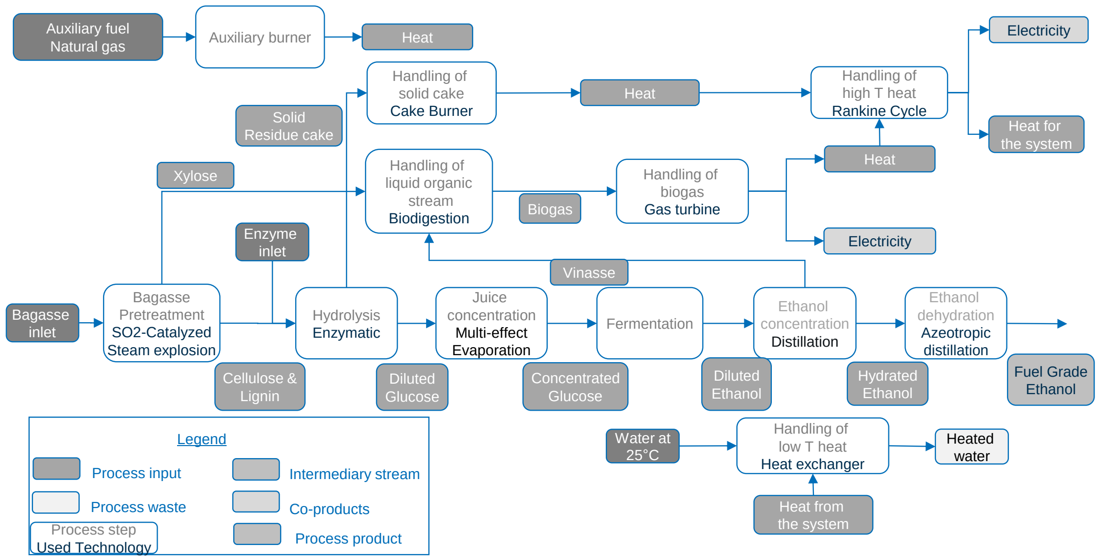

Figure 1: Process Layout with inputs, main steps, intermediary streams and products

3.1 Results for the base configuration

We hence launched a simulation with the annex HEN design model for our base case scenario. In Figure 2, we visualize the integrated composite curve for the auxiliary burner against the rest of the system. We can also visualize the heat duties of the various process sections. The separation section denotes the juice concentration, ethanol concentration and ethanol dehydration steps. As can be seen, there is a need for the auxiliary burner, and steam production network has not been activated. This is due to the great energy consumption of the system. The results for the parameters mentioned in Eq(1) are the following : W = -5.65 MW, $E_{cold\_utility}$ = -74 MW, $E_{auxiliary\_fuel}$ = -33 MW. This translates into a value for our objective function : U_{Prod}=-112.65 MW , a rather negative value. The negative value for the net power is due to the electricity consumption at the process level. The utility consumption occurs predominantly at the steam explosion and separation sections. It is on these sections that we will bring our attention in the optimisation phase.

4. Optimisation of net utility production

In this section, we seek to control and vary a given set of decision variables with the objective of optimising the net production of utility, U_{Prod}. These variables do not have a linear effect on the chosen objective function. For this reason, we make use of a nonlinear optimiser based on evolutionary programming, for this optimisation step (Ensinas et al., 2013). This optimiser is denoted hereafter as the Master Optimiser. The Master Optimiser sends a set of values for the decision variables to the process simulation model. For each set, the process is simulated, the HEN Design Model resolved and the objective function calculated. The optimiser behaves in a generational evolutionary manner : the system converges over the course of generations to the optimal solution. A key point at this level is hence the specification of the chosen design variables and their ranges. As specified earlier, these variables will pertain to the pretreatment and separation sections. The list of these variables, accompanied with a description of their effect, and their variation range is specified in Table 2.

Moreover some of these variables obey to a mass balance equation highlighted in Eq(2). This equation indicates that regardless of the fraction of water evaporated at a given effect, the sum of these fractions needs to be equal to one always.

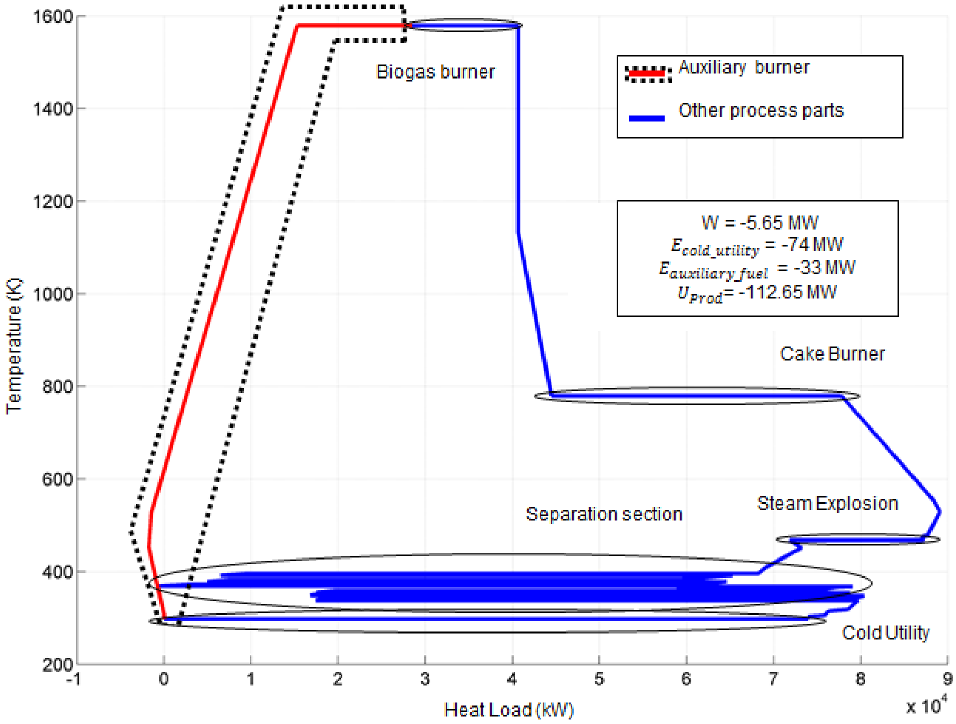

Figure 2: Integrated Composite Curve for auxiliary burner in base case

5. Description of the optimisation results

We hence performed a mono-objective optimisation, with 80 individuals per generation. The problem converged towards a single optimal point. The optimisation run was halted at the 47th generation, this due to the stagnation of the mean and standard deviation of the objective function and decision variables. The values for the decision variables, along with default values, are highlighted in Table 3. These values yield the following results for utility parameters : W = 11.28 MW (16.93 MW increase), $E_{cold\_utility}$ = -36.4 MW (36.6 MW increase), $E_{auxiliary\_fuel}$ = 0 MW (33 MW increase), U_{Prod} = -25.12 (86.53 MW increase). As we can see, there is a net increase in the value for the objective function This is due to a reduction in the system's energy consumption. This reduction is highlighted in the integrated composite curve of Figure 4. On this figure we visualize the absence of the auxiliary burner, as well as the activation of the steam network. Compared to Figure 2, the heat demand for the separation section went from 80 MW to 5 MW, and that for the steam explosion section went from 15 MW to 5 MW. Moreover we can visualize on Figure 4 the extent of the Heat Integration in the separation section.

Table 2: List of decision variables to be controlled in the optimisation section

Decision variable	Description	Effect on the system	Variation range	Default Value
S	Solids loading in the hydrolysis reactor	Effects the glucose concentration at the end of the hydrolysis step, Effects the heat demands of the separation section.	[2 %; 20 %]	2 %
$T_{pretreat}$	Temperature of steam explosion	Influences the steam consumption rate and subsequently the required evaporation heat	[180;205] °C	190 °C
∂_{ev}	Fraction of post-hydrolysis water discarded in the glucose concentration section.	Influences the integration between the ethanol concentration and glucose concentration sections	[20 %; 90 %]	95.45 %
$\partial_{ev,i}$	Fraction of water evaporated at a given effect. Six variables for six effects.	Influences the integration of the evaporator effect with the rest of the process	[0 %; 100 %]	16.7 %
∂_{dis_strip}	Fraction of post fermentation water discarded in the stripping column	Influence the energy integration of the distillation columns with the rest of the process	[50 %; 99.5 %]	96 %.
P_{dis_strip}	Stripping column pressure		[1;5] atm	1 atm
P_{dis_rec}	Rectification column pressure		[1;5] atm	1 atm

Table 3: Values for decision variables : Optimal value vs. Default Value

Decision Variable	S (%)	$T_{pretreat}$(°C)	∂_{ev} (%)	$\partial_{ev,1}$ (%)	$\partial_{ev,2}$ (%)	$\partial_{ev,3}$ (%)
Specified Value	19.9	180.03	46.7	0	29.4	0
Default Value	2	190	95.45	16.7	16.7	16.7
Decision Variable	$\partial_{ev,4}$ (%)	$\partial_{ev,5}$ (%)	$\partial_{ev,6}$ (%)	∂_{dis_split} (%)	$P_{dis,strip}$ (atm)	$P_{dis,rec}$ (atm)
Specified Value	18.9	20.9	30.8	99.3	1	1
Default Value	16.7	16.7	16.7	96	1	1

6. Conclusion and Perspectives

In this article, we highlighted a methodology for maximising the net utility production in a second generation ethanol production process, from bagasse. This net utility production comprises of net power production, heat lost to cold utility, and heat content of required hot utility. The methodology is based on an analysis of the process's composite curves. Twelve process variables were, controlled, via an evolutionary algorithm, to achieve the desired optimisation. As a result it proved more interesting to work: at a high solids loading in the hydrolysis reactor, 20 % in the optimal case, versus 2 % in the base case, at a low pretreatment temperature, 180 °C in the optimal case versus 190 °C in the base case, and with an enhanced integration between the separation sections On another hand, the net production of utility went from an initial value of -112.65 MW to an optimal value of -25.12 MW, an increase of 77.7 %. This methodology expands on previous research work namely (Ensinas et al., 2013), by providing a more solid simulation model capable of converging for multiple design configurations. This enabled a detailed study of the heat integration possibilities between the main separation sections and between the separation and reaction sections. However, at this point, certain design issues mentioned in (Dias et al., 2014) were not considered at this point. (1) The impact of the solids loading and the pre-treatment temperature on the hydrolysis reaction was not taken into account. In fact, the increase of the first parameter and the decrease of the second will reduce the reaction yield (Carrasco et al. 2010). (2) The increase in solids loading, will lead to a novel design for the hydrolysis reactor (Zhang et al. 2009). (3) Electricity consumption in the hydrolysis reactor was kept constant. (4) No alternative technologies for the different process blocks were taken into account (5) The economic and environmental impacts of the proposed alternatives were not assessed. The inclusion of these issues will lead to a more detailed application of the proposed methodology. It will also inevitably lead to a multi-route multi-objective problem, an issue that has not yet been wholly investigated in the case of ethanol production in literature.

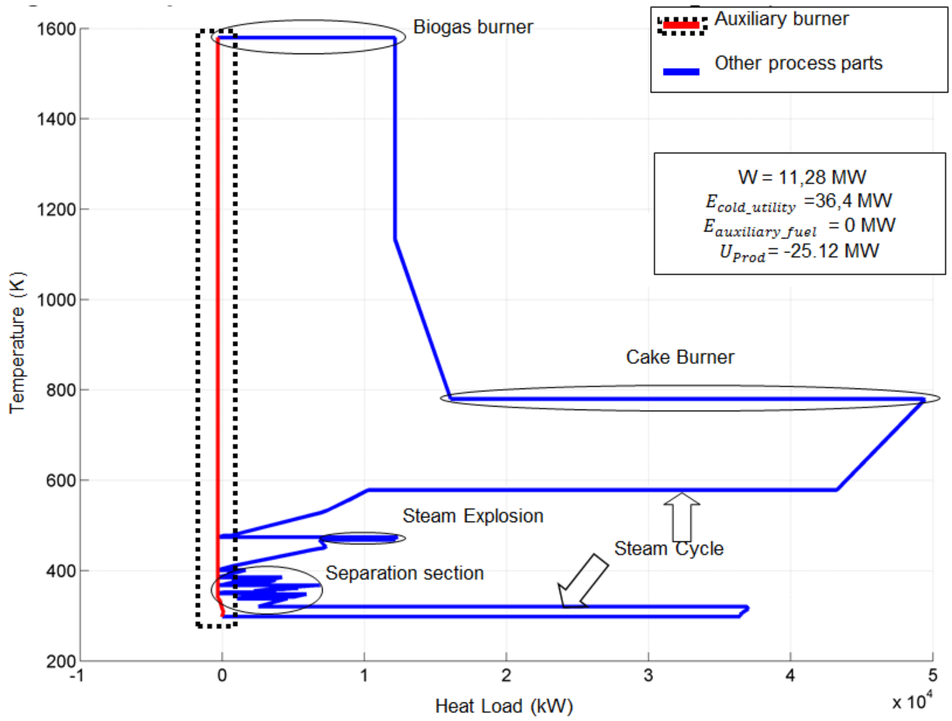

Figure 3: Integrated Composite Curve for auxiliary burner in optimal case

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