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Heat integration of combined 1st and 2nd generation ethanol production from wheat kernels and wheat straw

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Abstract

Background: Reducing the energy used in an ethanol plant is an important step towards reducing both the cost and environmental impact of the process. Pinch technology was used to analyze the energy utilization and to investigate possible energy savings in stand-alone and combined 1st and 2nd generation ethanol production plant designs. Different heat sources and sinks in the plant were identified to improve energy integration. Four different scenarios were evaluated using Aspen Plus, with heat exchanger networks generated in the simulation program Aspen Energy Analyzer. The total direct cost of the heat exchanger networks was calculated using Aspen Process Economic Analyzer.

Results: It was shown that heating costs could be reduced by 40-47% and cooling costs by 42-54% by heat integration. The sum of the discounted total direct cost for heat exchangers and annual cost of utilities was also lower in the heat integrated cases than in the corresponding non-heat integrated cases for all the configurations investigated.

Conclusion: Heat integration showed that the heating and cooling energy demands could be reduced to a great extent in stand-alone as well as combined 1st and 2nd generation bioethanol plants. The cost for heating and cooling of the process can be decreased with heat integration. The main cost for providing the processes with heat can be attributed to the cost of hot utilities.

Keywords: Heat integration, Bioethanol, First and second generation, Simulation, Wheat, Lignocellulosic, Pinch analysis

Background

Energy is required to secure and improve the production of many necessary commodities as well as to cover essential human needs, such as food and heating. The increasing global energy demand, combined with the effect of greenhouse gas (GHG) emission, which causes global warming, have accentuated the need to find sustainable and environmentally friendly energy sources. In the search for energy sources that can complement and replace fossil fuels it is important to look for alternatives that can meet these requirements. During the past decade, fossil fuels, which release high amounts of GHGs when burned, accounted for about 85% of the primary energy use in the World [1].

It has been estimated that in the USA, 28% of all GHG emission results from the transport sector, while industry accounts for around 20%, primarily originating from the use of fossil fuels [2]. The corresponding values for the EU are about 25% and 20%. Compared to the level in 1990, which was defined as the starting point for the EU's goal to decrease GHG emissions by 20% by 2020 [3], measurements made in 2007 showed that GHG emission resulting from transportation had increased by about 36%, while emissions from other sectors were generally decreasing. EU legislation has therefore required a further reduction of GHG emissions from vehicle fuels by up to 10% by 2020 as described in the amendment to the Fuel Quality Directive [4], which is closely related to the Renewable Energy Directive that is promoting 10% renewable energy in the transportation sector by 2020 [5]. Biofuels are seen as an important means of achieving the GHG emission target for the transport sector. The EU has also stated that it is imperative that fuels be produced in a

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sustainable way, in order to reduce the impact on the environment.

Bioethanol referred to as 1st generation (1G) ethanol is produced from food-grade material that often contains easily accessible sugars, such as sucrose from sugarcane, or starch from corn or wheat kernels. In order not to compete with food production, biofuels can be made from lignocellulosic residues such as agricultural residues and other lignocellulosic feedstock. Bioethanol produced from lignocellulosic material is often referred to as 2nd generation (2G) bioethanol, and in this case, the fermentable sugars are extracted from cellulose-containing materials. To date, only one commercial-scale facility is producing bioethanol from cellulose via the biochemical route. The plant, which is located in Crescentino, Italy, was constructed by Beta Renewables, and was opened at the beginning of October 2013 [6,7]. Some other commercial-scale facilities that are currently under construction, or planned for the near future are: the DuPont Biofuels Solutions' plant in NV, USA utilizing corn stover [8], Poet-DSM Advanced Biofuels' plant in Emmetsburg, IA, USA using corn cobs [9], Abengoa Bioenergy's plant in Hugoton, KS, USA using unspecified biomass [10], Project Alpha by Beta Renewables in Clinton, NC, USA using energy grass [11], Canergy's plant in The Imperial Valley of California, USA utilizing energy cane and straw [12], and the Iogen Corporation plant in São Paulo, Brazil using bagasse and straw [13]. To facilitate the introduction of and transition to 2G bioethanol, various strategies for the integration of 1G and 2G ethanol plants can be considered. Integration can help to reduce the cost of downstream processing in the 2G plant while the technology matures, and at the same time provide primary energy to the 1G plant. Wheat is an abundant crop in Europe today, which generates a large amount of straw. Bioethanol production from wheat kernels and wheat straw is thus of considerable interest.

Since the bioethanol production is considered as an energy demanding process it is important to keep track of the energy requirement in the plant. The energy utilization can be investigated using pinch analysis technology. Pinch technology was introduced by Linnhoff and Flower in 1978 [14], and is based on identifying heat sources and sinks in a process. These can then be used to provide the process with heating and cooling through the use of heat exchangers, reducing the need for external energy input. The cost of installing heat exchangers can then be compared to the cost of utilities in the form of external heating and cooling by steam and cooling water. Implementing pinch technology during process design can reduce investment costs during retrofitting [15].

A number of previous studies utilizing pinch technology have focused on improving the 1G bioethanol process to decrease the energy demand in the plant and to increase the surplus and use of residual material, excluding it from being incinerated. Dias et al. [16] investigated different methods of improving the process, such as increasing the degree of process integration in a sugar cane plant to reduce steam consumption, and thereby increase the amount of available lignocellulosic material. They also compared low- and high-pressure boiler systems to investigate a potential increase in electricity production. Their study showed that it was possible to increase the available amount of lignocellulosic material with increased process integration, which then can be used for bioethanol production in a 2G plant, or burned to generate electricity. It was also shown that a higher boiler pressure increased the electricity production but decreased the surplus of lignocellulosic material. The use of combined heat and power (CHP) generation, together with thermal integration of distillation or evaporation is often investigated during pinch analysis of a bioethanol plant, not only because distillation and evaporation are two of the most energy-demanding steps in the process, but also because optimizing the CHP process will result in overall better energy utilization. Morandin et al. [17] investigated a sugarcane-converting process with different configurations of a multi-effect evaporator and the option to integrate the process with a CHP plant fuelled by bagasse. By altering the design parameters for the evaporator they found that the hot utility demand could be reduced by one third, and that using bagasse to fuel the CHP plant could cover the energy demand for a base case scenario. Dias et al. [18] also demonstrated the benefits of incorporating distillation into the thermal analysis as this leads to a lower global heating demand for the production plant. Heat integration of the distillation and evaporation stages, at different pressure levels, has earlier also been modeled by e. g. Wingren et al. [19] and Sassner et al. [20] for the 2G ethanol plant leading to a decreased use of primary steam in the process. Pfeffer et al. [21] applied pinch analysis to a 1G bioethanol and a dried distillers grain with solubles (DDGS) plant utilizing wheat by generating a heat exchanger network, and found that heat integration of the beer column and the rectification column resulted in the highest primary heat reduction. The inclusion of a gas boiler or a gas engine in a CHP system was also simulated. Biogas was then produced instead of DDGS and used to replace external fuel. Franceschin et al. [22] used pinch analysis to optimize the energy and water demand in a dry-mill bioethanol corn plant. The results indicated that the water requirement could be reduced by implementing heat exchangers. Čuček et al. [23] employed pinch analysis to evaluate two different paths for syngas conversion of corn stover to bioethanol. The thermo-biochemical (syngas fermentation) and thermo-chemical (catalytic mixed-alcohol synthesis) processes were compared, and the latter showed better potential for heat integration, and therefore were preferred when

investigating integration of 1G and 2G bioethanol plants using corn and corn stover.

Some pinch analysis studies have also been performed on 2G stand-alone plants. Modarresi et al. [24] performed pinch and exergy analysis for a 2G wheat straw bioethanol plant with the aim of reducing the energy required for heating and cooling by implementing a heat exchanger network. In a another study, Kravanja et al. [25] investigated a distillation system containing two stripper columns and a multi-effect evaporator for a bioethanol process fermenting hexoses and pentoses from wheat straw. They found that one of the stripper columns and one evaporation unit could be removed by modifying the process design and implementing heat integration between the distillation and evaporation steps. Franceschin et al. [26] calculated the theoretical energy savings for a 2G bioethanol plant utilizing rye straw with co-production of xylitol and found it to be 68%. However, theoretical energy savings usually differ from the savings that are possible in a heat exchanger network in a real process. Two different studies investigating heat integration in bioethanol production from wood were performed by Piccolo and Bezzo [27] and Fujimoto et al. [28]. Piccolo and Bezzo [27] analyzed two different process alternatives (i.e. enzymatic hydrolysis and fermentation and gasification and fermentation) for production of bioethanol from lignocellulosic material and Fujimoto et al. [28] studied heat released in the process at low temperature. Fujimoto et al. [28] found that energy utilization was improved by introducing a mechanical vapor recompression (MVR) in the process. The MVR was used to increase the pressure of the overhead vapor from evaporation, which could be used later as a source of heat for distillation and drying. The use of MVR in the evaporation step for a 2G plant has also been studied by Sassner et al. [20].

In the present study, pinch analysis was applied to evaluate various cases consisting of 1G and 2G stand-alone plants, co-located 1G and 2G plants and a combined 1G and 2G plant in which material streams are mixed prior to distillation. The study was performed to investigate the effect of employing heat exchanger networks in full-scale plants. The simulation program Aspen Energy Analyzer was used to create heat exchanger networks from process models generated in the flowsheeting program Aspen Plus. The total direct cost of the heat exchanger networks was subsequently estimated using Aspen Process Economic Analyzer and compared to the cost of utilities.

Results and discussion
Energy
The composite curves that were obtained from the pinch analysis for the four different base cases, using a global minimum ΔT of 10°C, are shown in Figures 1, 2, 3 and 4. Preheating of the incoming wheat and straw material, which had a high solids content, was performed with steam or condensate. Steam-pretreated material was cooled with condensate or cooling utility. In some case, for example, for condensing steam, a lower value of ΔT could be used since the heat transfer coefficients of condensing steam are higher than for liquids. This can result in a lower usage of heating and cooling in the heat integrated cases compared with the minimum heating and cooling target that can be subtracted from the composite curves in Figures 1, 2, 3 and 4.

The hot and cold pinch temperatures appeared at 113°C and 103°C, respectively, for the 1G stand-alone plant and the co-located plants. For the 2G stand-alone plant the hot pinch appeared at 95°C and the cold pinch at 85°C. For the combined case the corresponding values were
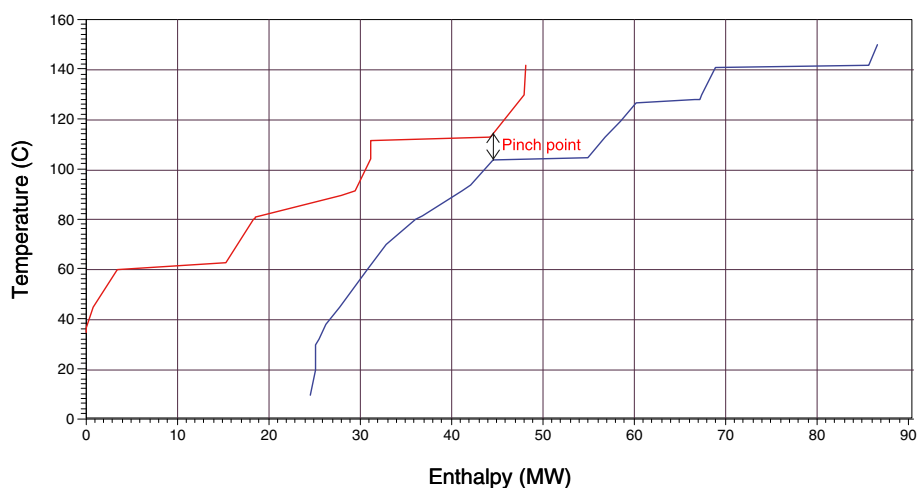

Figure 1 Hot and cold composite curves for the 1G ethanol plant. Resulting in a minimum heating and cooling target of 39 MW and 25 MW, respectively.
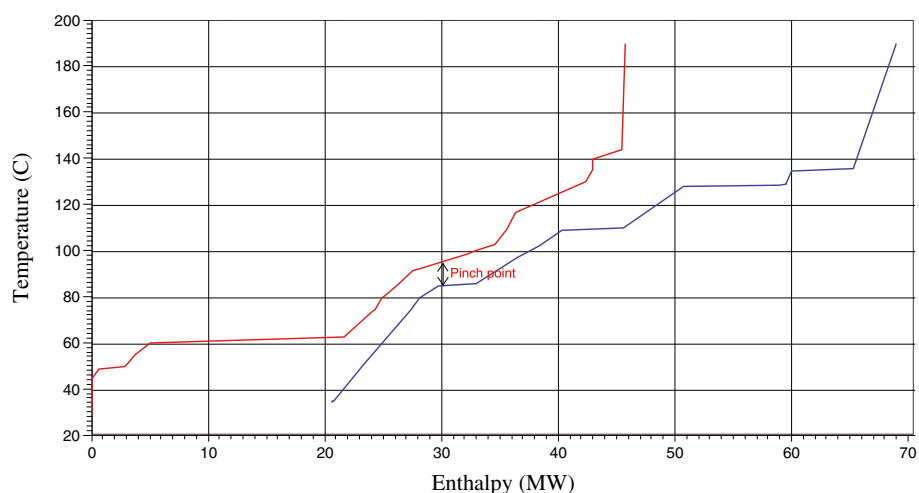
Figure 2 Hot and cold composite curves for the 2G ethanol plant. Resulting in a minimum heating and cooling target of 23 MW and 21 MW, respectively.

93°C and 83°C. The approximate total heat transfer area and number of heat exchangers for the different non-heat integrated and heat integrated cases are also presented in Table 1. It shows that the total heat transfer area and numbers of heat exchangers in all cases are lower for the non-heat integrated cases. The large difference in heat exchanger area in the non-heat integrated and heat integrated cases can be explained by the larger ΔT that will originate when utility streams are utilized for heat transfer instead of process streams. A larger ΔT will result in a lower heat transfer area. In the table it can also be seen that it is a small difference in the number of heat exchangers between the non-integrated and integrated case. This can occur since all streams will need to be heat exchanged at least once, both in the non-heat integrated and

the heat integrated cases. A restriction was also set on the maximum amount of stream splitting that was allowed to be made.

The heating and the cooling demands for the four base cases, each case evaluated with and without heat integration, are presented in Table 2. Heat sources and sinks in the plants were covered either solely by utilities or both utilities and heat integration, by heat exchange. For the 1G integrated case it can be seen that the heating and cooling demands will be less than the minimum heating and cooling demands presented in Figure 1. This results from that a lower ΔT than the global ΔT minimum sometimes can be utilized in a single heat exchanger, potentially leading to a lower overall heating or cooling demand. In this case, the reboiler of the stripper, for example, was heat
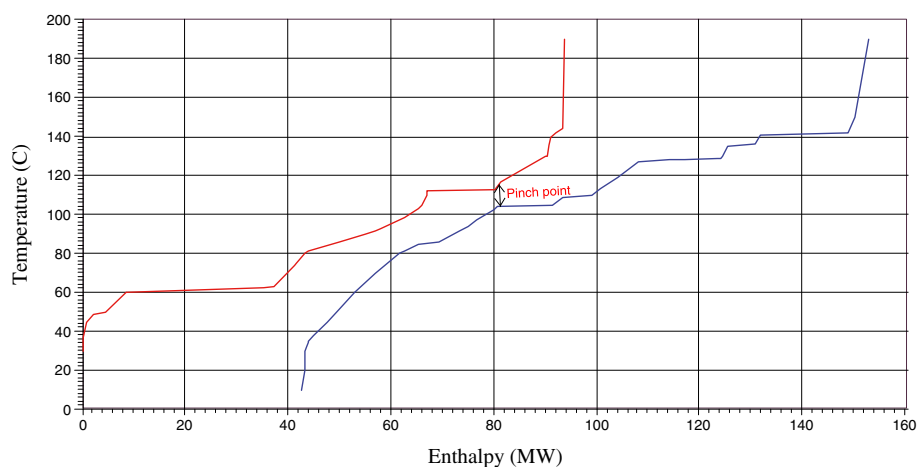

Figure 3 Hot and cold composite curves for the co-located 1G and 2G ethanol plants. Resulting in a minimum heating and cooling target of 59 MW and 43 MW, respectively.
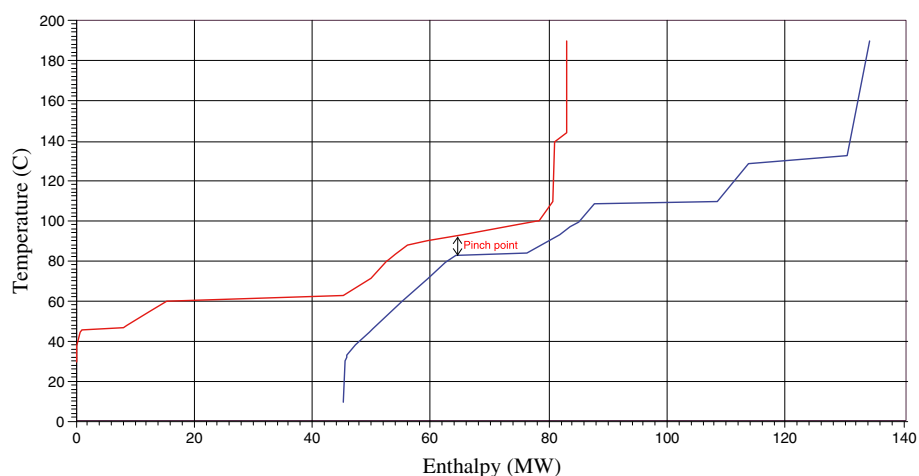
Figure 4 Hot and cold composite curves for the combined 1G and 2G ethanol plant. Resulting in a minimum heating and cooling target of 51 MW and 45 MW, respectively.

exchanged with the overhead vapor from the rectifier, which resulted in a lower ΔT than the global ΔT. This was assumed possible since the heat transfer coefficient for the condensing steam is higher than for a liquid and consequently a lower ΔT can be appointed over the heat exchanger. Table 2 also shows the energy required to produce one kg of ethanol in each case. An electricity consumption of 0.84 MJ/kg ethanol was also added to each case when calculating the energy demand per kg ethanol, as this value corresponds to the average electricity demand in a 1G corn ethanol plant [29].

The potential energy savings that can be made according to the pinch analysis for the various cases studied are presented in Figure 5. The whole bars (100%) in Figure 5 represent the hot and cold utility requirements for the non-heat integrated plants. The top parts of the bars represent the savings in heating and cooling, which could be obtained by implementing heat integration, compared with the plants without heat integration, and the bottom parts the energy utilized.

The results shown in Figure 5 indicate that savings of between 40 and 47% could be made in heating, and

between 42 and 54% in cooling, depending on the plant configuration, by implementing heat integration. This illustrates the considerable potential for major energy savings when designing new 2G or combined 1G and 2G ethanol plants. Considerable heat exchange is already being employed in 1G plants, but further improvements may be possible. As the various cases represent different raw materials, loadings and plant configurations, the results should primarily be regarded in relation to each individual base case. Also as there are several ways of constructing an energy-efficient heat exchanger network, depending on the boundary conditions, these results are only some of a number of possible solutions. The optimal solution will depend on the kind of process in question, the desired end products, and the feasibility of the modeled network, however some conclusions can be drawn.

It can be seen from Table 2 that the energy demand of the combined plant (88 MW) is lower than that of the co-located plants (108 MW) using the same raw material loading, and thus producing the same amount of ethanol. When heat integration is implemented the combined case requires 53 MW heat, while the co-located plants use 63 MW. The combined cases thus requires 16-19% less energy than the co-located cases. The improvement could be partly explained by the process design. For example, no dryer is needed in the combined cases since no DDGS are produced. Moreover, only one evaporation and one distillation unit are needed in the combined cases since the 1G and 2G streams are combined prior to distillation. When no DDGS are produced all the residual material after filtration and evaporation is incinerated together with the biogas produced to supply the plant with heat. Any excess heat produced during the incineration, assuming the demand of the plant is met, will be used to generate electricity that can be sold.

Table 1 Total heat transfer area and number of heat exchangers in the non-heat integrated and heat integrated cases

	1G	2G	Co-located 1G & 2G	Combined 1G & 2G
Heat integrated				
Total heat transfer area (m²)	32000	25500	56300	48200
Number of heat exchangers	28	30	56	29
Non-heat integrated				
Total heat transfer area (m²)	14400	14900	29300	23500
Number of heat exchangers	24	27	51	28

Table 2 Heating and cooling demands for the four ethanol plant investigated, with and without heat integration

	1G		2G		Co-located 1G & 2G		Combined 1G & 2G	
	Hot	Cold	Hot	Cold	Hot	Cold	Hot	Cold
Heating and cooling demand								
Without heat integration (MW)	63	49	48	46	108	95	88	83
With heat integration (MW)	36	22	25	23	63	47	53	48
Energy demand per kg ethanol produced								
Without heat integration (MJ/kg)	25	20	42	40	31	28	26	25
With heat integration (MJ/kg)	15	10	23	21	19	15	17	16

The energy used per kg ethanol produced in the process is presented in the last two lines.

This can be seen in Table 3 where the combined plant with heat integration will produce 31 MW of electricity, which can be compared with the co-located plants that produce 17 MW of electricity. Depending on the market for DDGS and electricity, the co-located heat integrated plants or the combined heat integrated plant is more likely to be feasible. In Table 2 it can also be seen that the combined plant with heat integration (53 MW) requires only 49% of the energy used by the co-located plants without heat integration (108 MW).

The results obtained for the 2G stand-alone ethanol plant indicate that the savings in the energy required for heating by utilizing an optimized heat exchanger network could be up to 47%, without any other changes in the process. Kravanja et al. [25] have studied how design modification of the distillation and evaporation stages and integration with the background process could affect the energy demand, indicating that it could be decreased even further.

Table 2 also gives the energy required to produce one kg of ethanol in each case. The calculations are based on

the energy utilization (including external electricity) of each plant and the corresponding ethanol production, which varied depending on the raw material loading. The highest heating energy demand was found for the 2G stand-alone plant, and the lowest for the 1G stand-alone plant, both with and without heat integration. The 2G plant without heat integration used 42 MJ/kg ethanol, while the case with heat integration required only 23 MJ/kg ethanol. The corresponding values for the 1G plant were 25 MJ/kg ethanol and 15 MJ/kg ethanol. The results obtained for the 2G plant are in line with the savings found in a previous study of a similar 2G ethanol plant by Modarresi et al. [24], who obtained values of about 32 MJ/kg ethanol for a plant without heat integration and 19 MJ/kg ethanol for one with heat integration. The presented results for the 1G plant can be compared with results found in the literature for a 1G corn ethanol plant utilizing approximately 27 MJ/kg ethanol in a non-optimized and 14 MJ/kg in a more efficient plant configuration [29]. Pfeffer et al. [21] also calculated that with a ΔT of 20°C a feasible wheat bioethanol plant producing
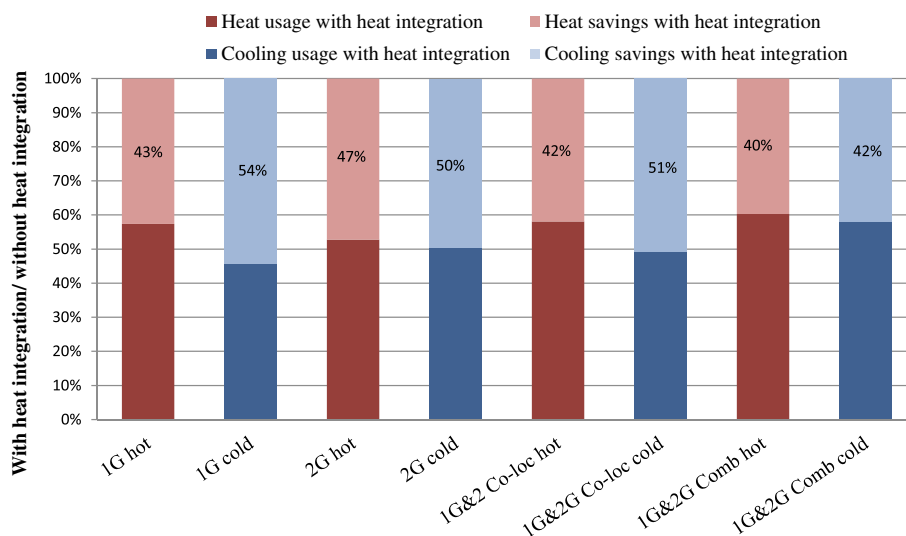

Figure 5 Percentage of energy usage and potential savings for hot and cold utilities. Comparing heat integrated cases with non-heat integrated cases.

Table 3 Energy inflow of raw materials and surplus heat and energy outflow of products including the mass flow of DDGS

	1G	2G	Co-located 1G & 2G	Combined 1G & 2G
Energy inflow				
Raw material (MW)	136	119	256	256
Natural gas (methane) added: with heat integration (MW)	37	0	11	0
Natural gas (methane) added: without heat integration (MW)	64	0	60	0
Electricity (MW)	2	1	3	3
Products				
Total biogas production: with heat integration (MW)	0	7.5	7.5	8.4
Total biogas production: without heat integration (MW)	0	7.5	7.5	8.4
Biogas surplus: with heat integration (MW)	0	7.5	0.0	8.4
Biogas surplus: without heat integration (MW)	0	6.8	0.0	0.5
Ethanol production (MW)	75	30	105	106
Total electricity production: with heat integration (MW)	0	18	17	31
Total electricity production: without heat integration (MW)	0	15	17	27
DDGS production (kg/h)	14134	0	14134	0

DDGS would utilize around 20 MJ/kg ethanol, excluding electricity.

A lower heat demand per kg ethanol was required in the heat integrated cases by both the co-located plants (19 MJ/kg ethanol) and the combined plant (17 MJ/kg ethanol), than by the 2G stand-alone plant with heat integration (23 MJ/kg ethanol). Comparison of the 1G stand-alone plant and the combined plant reveals that the values are similar, namely 17 MJ/kg ethanol and 15 MJ/kg ethanol, respectively.

The main heating energy requirements in the base cases are those for the distillation- and evaporation stages, the preheating of the distillation- and evaporation feed and the dryer. The distillation units use between 28% (the 2G stand-alone plant) and 43% (the 1G stand-alone plant) of the total energy required by the base cases, without heat integration. Depending on the pressure in the distillation columns, and the size of the feed flow, the energy required to preheat the feed to the distillation stage will vary between 11%, for the 1G stand-alone plant, and 28%, for the 2G stand-alone plant. In the 1G plant the feed is heated to a lower temperature than in the 2G plant since the pressure in the columns is operated from low to high pressure so as to avoid gluten deposits, while the columns in the 2G plant are operated from high to low pressure. The evaporation units consume between 11 and 19% of the total energy usage for the various cases without heat integration, where the 1G plant requires the lowest, and the combined plant and the 2G stand-alone plant the highest, since the stillage is more diluted in these cases. Preheating the feed prior to evaporation in the non-heat integrated cases would require between 7 and 14% of the total energy

used by the different plants. The 2G stand-alone plant and the combined plant requires a high energy input to preheat the evaporation feed, due to the expected low ethanol concentration in the distillation feed, resulting in a large stillage stream. Finally, the dryer utilized 10-17% of the total energy requirement in the 1G stand-alone and the co-located cases without heat integration.

Pinch analysis showed that the energy required for distillation could be reduced by 38-60%, depending on the case, when implementing heat integration. The lowest reduction, 38%, was found for the 1G stand-alone plant. This is mainly the result of the high temperature in the rectifying reboiler (143°C) preventing the transfer of heat from other streams in the process to the reboiler because of the lower temperatures in the rest of the process. The highest energy reduction in distillation, 60%, was found for the 2G stand-alone plant. Part of the energy demand was covered by utilizing condensing steam from the pretreatment step. The corresponding values for the co-located plant with heat integration and the combined plant with heat integration were 52% and 42%, respectively. The energy required to heat the distillation feed in the combined heat integrated plant was almost zero, and for the other cases the reductions were 64-81%. These reductions were partly the result of heat exchange between the stillage streams and the feed streams. The energy required to heat the feed stream to the evaporation unit was also greatly reduced for the 1G and 2G stand-alone plants with heat integration, being 82% and 90%, respectively. For the combined plant the reduction was only 8% after heat integration, and for the co-located plant it was about 30%. No heat reduction was observed in the evaporation stage. The reason for this was

mainly the high temperature of 129°C in the first effect, which limited the amount of available streams that could transfer energy to the evaporator stage. The condensing streams that could have been used for heat exchange in these cases were primarily chosen to supply the distillation stage with energy. Therefore, utilities had to be supplied to the evaporation stage. For the high-temperature condensing streams the primary options for heat exchange are in distillation, evaporation and in the dryer (but also, to some degree, the feed streams) since these steps are operated at the highest temperatures. Therefore, it is important to investigate the complete process to obtain a good overview of all the heat sources and heat sinks in the system. A large amount of energy must be dissipated at low temperature because of the low pressure in the last evaporation step, where the overhead vapor has a temperature of 63°C. This energy could partly be recovered with MVR, as investigated by Fujimoto et al. [28] and Sassner et al. [20]. Part of the steam could possibly also be used to replace some of the live steam in the liquefaction step.

Table 3 presents the energy inflow and the power and DDGS produced in the various cases investigated. The raw material loadings are presented as the heat of combustion of the raw material based on the lower heating values. Natural gas (methane) added in the table represents extra heat energy input needed to fuel the plants when residual material was not enough or non-existent to supply the plants with heat. Electricity listed in the table under energy inputs represents the assumed electricity requirement of 0.84 MJ/kg ethanol, corresponding to an average electricity demand in a 1G ethanol corn plant. Because of these assumptions the electricity listed in Table 3 is not subtracted from the electricity listed

under products, which is representing the total electricity generated in the various cases. Biogas and electricity were produced in all cases except the 1G plant, where DDGS were the only co-product. The biogas produced was primarily utilized to cover the energy demand in the plants, however, in some of the cases a surplus of biogas was generated; therefore, both the total produced biogas and the surplus biogas, not utilized in the plant, are included in the table.

With the plant capacities assumed in this study, only the 2G stand-alone plant and the combined plant are self-supporting in heating energy for the process, by burning residual material. This applies to the cases with and without heat integration. In all the other cases, natural gas (methane) had to be added to cover the energy demand of the process. Natural gas was used as the only energy source for the 1G plant since the DDGS produced was intended to be sold, and there was therefore no residual material that could be incinerated to supply the plant with heat. This shows that with the plant capacity chosen for the 2G plant the residual material produced in the 2G process is not sufficient to supply both the 1G and 2G stand-alone processes with heat. However, the use of external energy can be reduced to about one third in the 1G plant while still supplying the 2G plant with all the heat required. It should also be kept in mind that the 1G plant can be fueled with residual straw or other biomass instead of natural gas. Figure 6 shows the energy efficiency of the products (including DDGS) in relation to the energy input. The lower heating values (LHV): 50 MJ/kg for methane, 27.1 MJ/kg for ethanol, 17.1 MJ/kg for wheat straw, and 16.3 MJ/kg for wheat kernels were used. An average lower heating value of
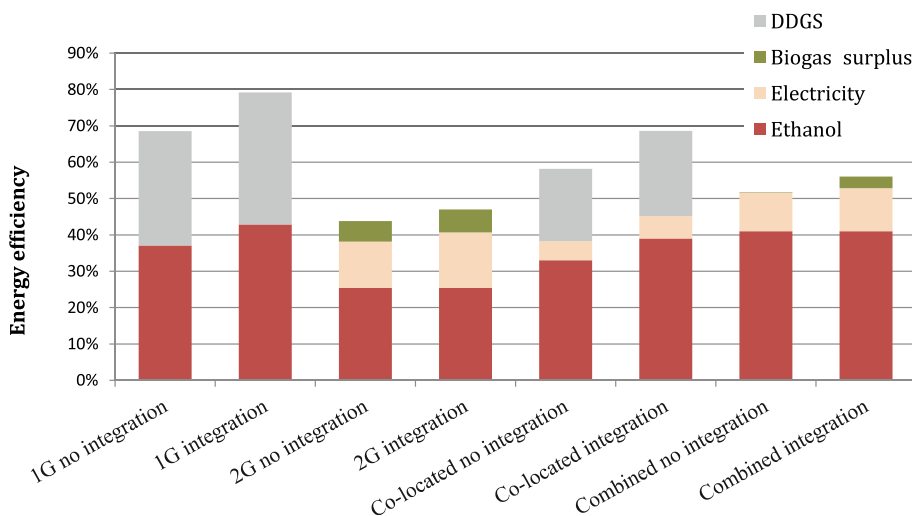

Figure 6 Energy efficiency on LHV basis for ethanol, electricity, biogas and DDGS related to energy inflow. The energy inflow includes raw material, electricity and natural gas for the various cases investigated.

Table 4 Cost of utilities and the total direct cost of heat exchangers per annum

	1G	2G	Co-located 1G & 2G	Combined 1G & 2G
Heat integrated				
Steam: with heat integration (MEUR/a)	10	8	18	15
Water: with heat integration (MEUR/a)	0.27	0.28	0.57	0.59
Heat exchangers: with heat integration (MEUR/a)	2.41	2.04	4.32	4.12
Total cost of utilities and heat exchangers: with heat integration (MEUR/a)	**13**	**10**	**23**	**20**
Non-heat integrated				
Steam: without heat integration (MEUR/a)	17	14	30	24
Water: without heat integration (MEUR/a)	1	1	1	1
Heat exchangers: without heat integration (MEUR/a)	1.25	1.29	2.54	2.05
Total cost of utilities and heat exchangers: without heat integration (MEUR/a)	**19**	**16**	**34**	**27**

18.7 MJ/kg dry matter was used for the DDGS [30,31]. Energy input included the raw material, supplemented methane and an average electricity requirement of 0.84 MJ/kg ethanol. The energy efficiency for the DDGS is included in Figure 6 to show the total energy output of the co-products. However, the DDGS were not considered for incineration in these cases, and their energy efficiency is, therefore, not further discussed.

In all cases where natural gas must be added the energy efficiency will decrease. The results show that the combined plant had a higher energy efficiency for ethanol and electricity than the co-located plants at the same raw material loading, both with and without heat integration. However, DDGS is also produced in the co-located plants, and the market price of these should also be considered.

Economics

Table 4 presents the total annual cost of utilities (steam and cooling water) in million EUR (MEUR) for the different cases. Direct steam used for steam explosion and liquefaction is also included. The total direct cost of heat exchangers, calculated with APEA, and given in the table, includes an annuity of 0.11, corresponding to 7% internal rate, and an economic lifetime of 15 years.

The results presented in Table 4 show that the cost of heat exchangers is higher for the heat integrated cases than in the non-heat integrated cases, however the cost of utilities is lower. Figure 7 shows the annual total cost of hot and cold utilities and heat exchangers for each case.

It can be seen that the total cost is lower in all cases with heat integration than in the corresponding cases without heat integration. The cases with heat integration
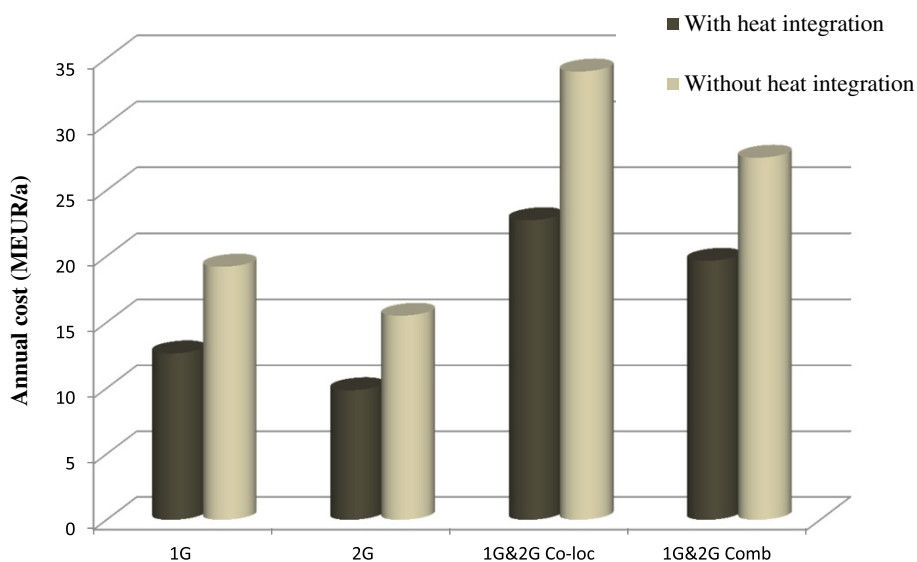

Figure 7 Sum of the discounted total direct cost for heat exchangers and annual cost of utilities (MEUR: EUR millions).

are 28-37% less expensive than the cases without heat integration, depending on the case. The cost of utilities is the main cost item, and corresponds to 75-90% of the total cost for the heat integrated and non-heat integrated cases, depending on the case. About 90% of the cost of utilities is for heating. In the cases including heat integration, the total cost of utilities and total direct costs for heat exchangers will be between 13 and 23 million EUR per year.

A sensitivity analysis was also carried out to investigate the annual total cost of utility and discounted total direct cost of heat exchangers when the cost of the hot utility was decreased, compared to the base case, assuming a base case steam cost of 20 EUR/ton for 4- and 6-bar steam and 25 EUR/ton for 20-bar steam. The result of the sensitivity analysis is presented in Figure 8.

Figure 8 shows that the price of steam must be reduced to less than about 25% of the original price to make the non-heat integrated cases less costly than those heat integrated.

Conclusions

Pinch analysis was applied to a selection of cases: a 1G stand-alone plant, a 2G stand-alone plant, co-located 1G and 2G stand-alone plants and a combined 1G and 2G plant, showing that savings of 40-47% could be made in hot utilities and 42-54% in cold utilities by heat integration.

For all cases it can clearly been seen that the total cost, in this case including the discounted total direct cost of heat exchangers and the utilities, is lower in the cases with heat integration than in the corresponding cases without heat integration, the former being between 28

and 37% less expensive than the latter. The main cost item is the cost of utilities, which accounts for more than 75-90% of the total cost, depending on the case. Approximately 90% of the utility cost was for heating.

The energy requirement (electricity included) of producing 1 kg of ethanol was lower both for the heat integrated 1G and 2G configuration (17 MJ/kg ethanol), and the co-located 1G and 2G stand-alone plants (19 MJ/kg ethanol) compared with the 2G stand-alone plant (23 MJ/kg). The energy demand of producing 1 kg of ethanol with the heat integrated 1G and 2G plant was similar to that of a 1G heat integrated stand-alone plant, which was 15 MJ/kg.

It was also shown that with the plant capacities assumed here, only the 2G stand-alone plant and the combined 1G and 2G plant were self-sufficient in heat by burning the residual material and biogas produced in the process. In the other cases external energy had to be added to supply the process with heat. Heat had to be supplied to the 1G plant as it was assumed that the DDGS produced would be sold, and thus no residual material would be available for incineration. The residual material obtained from the 2G process was not enough to supply both the 1G and 2G plants with energy. However, a reduction of about 30% in the external energy required for heating in the 1G plant could be accomplished, while completely satisfying the heat demand in the 2G plant.

Future work

The energy efficiency of the ethanol production process is an ongoing challenge. The flue gas, which leaves the CHP plant at 150°C, can be used to supply the dryer with heat, reducing the hot air needed in the drying process. MVR
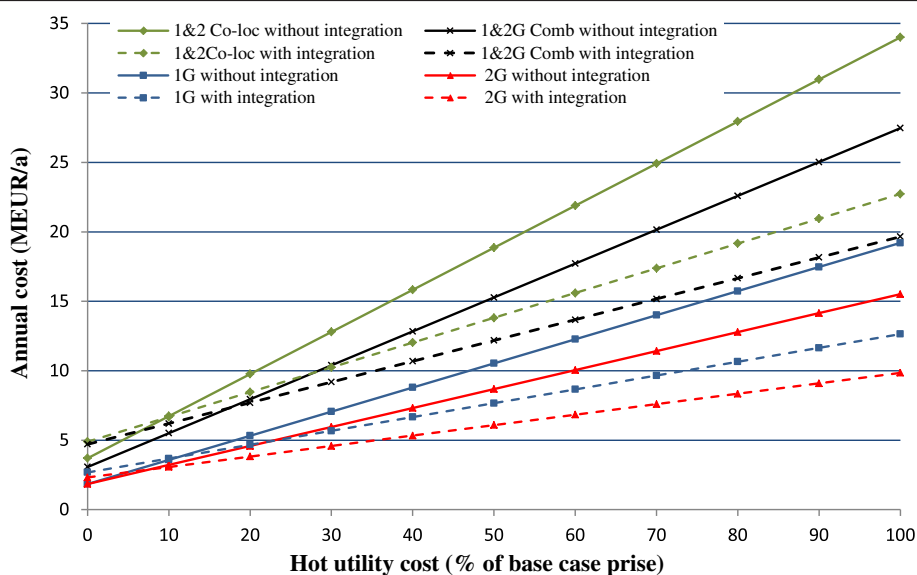

Figure 8 Decreased steam cost. The annual total cost of utilities and discounted total direct cost for heat exchangers when the cost of steam was decreased compared with the base case prices.

can be used to increase the pressure of the overhead vapor leaving the evaporation stage. This vapor could then potentially replace some of the live steam in the liquefaction step. Investigations on extracting the distillation, evaporation and the pretreatment stages from the rest of the streams in the background process are also recommended to identify possible additional savings in heating and cooling demand. The ratio of raw material loading between the 1G and 2G plant is a crucial factor, for the latter one to being able to supply the 1G plant with heat. The implementation of pinch analysis by connecting a heat integrated bioethanol and CHP plant to a local district heating network, may improve the heat utilization but a techno-economical evaluation which can verify the cost effectiveness should be performed.

Materials and methods
Process model
A facility with the potential to include both a 1G and 2G ethanol plant operating for 8000 hours per year was considered. An annual ethanol production of approximately 142000 m^3 was assumed for the 1G and 2G plants. This would corresponds to a 1G raw material loading of 30 ton dry wheat kernels per hour and 2G raw material loading of 25 ton dry wheat straw per hour assuming an ethanol production of about 101000 m^3/annum for the 1G plant and 41000 m^3/annum for the 2G plant.

Simulations for 1G and 2G ethanol plant configurations were carried out using the simulation program Aspen Plus, implementing experimental data from lab-scale work performed by Linde et al. [32] and Erdei et al. [33], and the process development unit at the Department of Chemical Engineering, Lund University. The models for the 2G plant used in this study were modified versions of previous models developed by Wingren et al. [19,34] and Sassner et al. [35]. One of the major differences in the current model compared with that studied by Sassner et al. [35] is the choice of different materials; in this case wheat straw, instead of wood. Therefore, the yield and design were adjusted to the use of wheat straw. The production of lignin pellets was not either considered in this study; instead the residual material produced in the 2G process, including lignin, was primarily intended for combustion. In some of the configurations where 1G ethanol production was included, DDGS were also produced, and it was assumed that these would be sold for use as animal feed. Lab-scale data were used to model the pretreatment and the simultaneous saccharification and fermentation (SSF) steps in the simulation of a 2G plant.

Simulations were carried out for a 1G stand-alone plant, for a 2G stand-alone plant, for co-located 1G and 2G stand-alone plants and for a 1G and 2G plant combined prior to distillation. Heat integration was implemented in these four cases and the cost of utilities and the total direct cost for heat exchangers were calculated. Non-heat integrated cases were chosen as references as no existing plant data are available for the 2G case to this date that could serve as a base case for the analysis. The reference cases will therefore indicate the least favorable energy utilization but they were chosen to facilitate the model comparison and illustrate the potential of energy integration. The process designs for the two stand-alone plants and the combined plants are described below. The co-located 1G and 2G case is simply a combination of the two stand-alone plants, and is therefore not described separately. The raw material loadings, surplus heat added, conversion factors in enzymatic hydrolysis and fermentation and products are listed in Table 5 for the various cases. The conversion factors and products are described more closely under each plant description section and the Results section, respectively. The surplus heat added, represented in the table as natural gas, is described further under the Energy supply section.

Description of the 1G stand-alone plant
Liquefaction
It was assumed that wheat kernels were mixed with a blend of fresh and recirculated water in a mixing tank to produce a mash that was preheated to 60°C. The mash was further heated by injecting live steam at 94°C. An enzyme mixture consisting of α-amylases was added to the mash and liquefaction was carried out at 90°C [36]. The mash was cooled to 37°C and then mixed with fresh water, cooling it further to 32°C before saccharification and fermentation. The process configuration for the 1G stand-alone ethanol plant is illustrated in Figure 9.

SSF in 1G
The cooled, diluted mash was fed into a fermentor where additional enzymes were added, and SSF was performed at 32°C. Water was added to achieve a final ethanol concentration of 8% by weight. The ethanol yield over fermentation for the 1G starch material was obtained in lab-scale experiments performed by Erdei et al. [33]. Based on the results, starch to glucose conversion after the enzymatic hydrolysis was set to 0.99 and the glucose-to-ethanol yield in fermentation was set to 0.94, referring to the starch content of the substrate. The starch to ethanol yield over the 1G SSF process will then be set to 0.93 as can be seen in Table 5.

Distillation in 1G
The 1G distillation process is assumed to consist of a 30-stage, low-pressure stripper column followed by a 60-stage, high-pressure rectifying column. The temperature in the stripper column is restricted to about 100°C due to the risk of gluten fouling in the column. The SSF broth is preheated to 91°C before entering the stripper column, operating at a top-stage pressure of 1 bar and a

Table 5 Plant capacity and conversion factors

	1G	2G	Co-located 1G & 2G	Combined 1G & 2G
Raw material and natural gas amount				
Wheat kernels (kg/h)	30000	-	30000	30000
Wheat straw (kg/h)	-	25000	25000	25000
Natural gas (methane) added: with heat integration (kg/h)	2634	0	808	0
Natural gas (methane) added: without heat integration (kg/h)	4593	0	4320	0
Conversions				
Starch to glucose conversion	0.99	-	0.99	0.99
Glucan to glucose conversion	-	0.91	0.91	0.91
Glucose to ethanol conversion in fermentation	0.94	0.92	1G: 0.94	1G: 0.94
			2G: 0.92	2G: 0.92
Ethanol yield after SSF	0.93	0.84	1G: 0.93	1G: 0.93
			2G: 0.84	2G: 0.84
Products				
Total biogas production (kg/h)	0	1086	1086	1208
Ethanol production (kg/h)	9956	4044	14000	14115
DDGS production (kg/h)	14134	0	14134	0
Total electricity production: with heat integration (MW)	0	18	17	31
Total electricity production: without heat integration (MW)	0	15	17	27

Murphree efficiency of 50%. Overhead vapor from the stripper containing an ethanol concentration of about 50% was condensed and preheated to 120°C before being fed to a rectifying column, which operates at a top-stage pressure of 3.5 bar, having a Murphree efficiency of 75%. The reflux ratio was shifting slightly between the non-heat integrated and heat integrated rectifying columns to a final value of about 4.2 and 3.9 for the non-heat integrated and heat integrated, respectively. The vapor exiting the rectifier, containing an ethanol concentration of 92.5%, is superheated to 130°C and then fed to a molecular sieve for further dehydration. The dehydrated ethanol stream is modeled as containing 99.5% ethanol. The rejected stream is condensed at 113°C and recirculated back to the rectifying column. The dehydrated ethanol is condensed and then cooled to a temperature of 35°C. Stillage from

the stripper column (thick stillage), containing liquids and solids, is cooled to 80°C before being transferred to a solid-liquid separation step. The rectifying stillage is cooled to 45°C and part of it is recirculated to the liquefaction step, while the remainder is sent to a waste-water treatment (WWT) unit.

Evaporation and solids separation in 1G

Solid-liquid separation is applied to the thick stillage, containing a dry matter of 11%, to remove most of the solids from the liquid to achieve a thin stillage. Separation is assumed to be carried out with a filter unit with a solids retention of about 95% [37]. The thin stillage and the wet cake have dry matter contents of approximately 5% and 60%, respectively. The thin stillage was heated to 128°C before being sent to the evaporator.
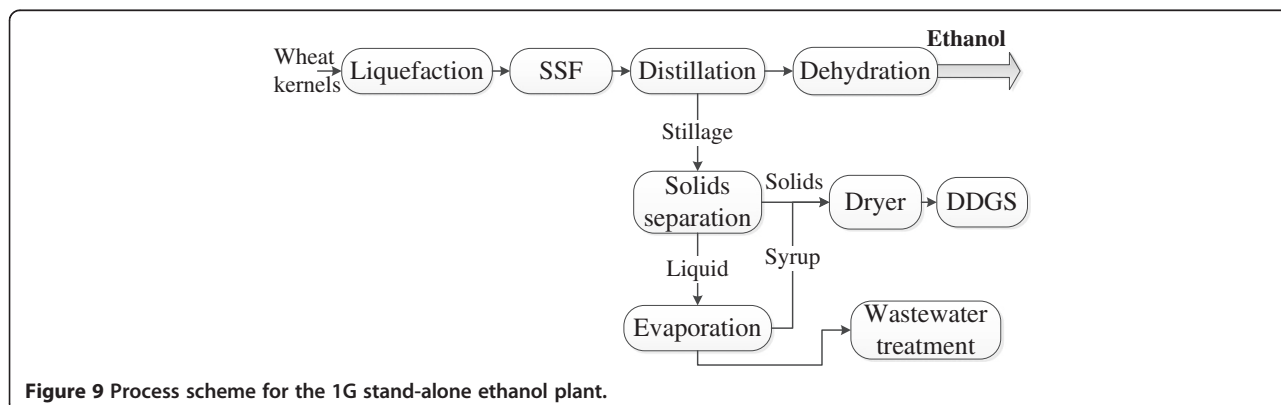

Figure 9 Process scheme for the 1G stand-alone ethanol plant.

The evaporation unit is modeled as a five-effect, forward-feed evaporation system. The boiling point elevation is modeled according to an expression presented by Larsson et al. [38]. The outgoing liquid fraction, referred to as the syrup, is assumed to contain 60% solids (dry matter) and to leave the evaporator at a pressure of 0.2 bar. The first-effect evaporator is supplied with steam at 6 bar as heating medium. The overhead vapor from the last effect of the evaporator is condensed at about 63°C, cooled to 45°C, and partly used as recirculation water; the excess being sent to the WWT unit. The condensate resulting from the five evaporation effects is likewise cooled to 45°C before being sent to the WWT.

Drying

After evaporation, the syrup and the wet cake are mixed and fed to an air dryer. The dryer is assumed to operate at an ingoing air temperature of 150°C and a discharge temperature of 90°C. Seventy percent of the outgoing air is recirculated and mixed with preheated fresh air. The incoming fresh air is assumed to have a water content of 0.005 kg H_2O/kg dry air, and is heated from 10°C to 20°C before being mixed. The air mixture is heated to 150°C before entering the dryer. The dry matter content of the outgoing DDGS is set to 88% [39].

Description of the 2G stand-alone plant
Pretreatment

Pretreatment was modeled based on lab-scale results obtained using wheat straw impregnated with H_2SO_4, which was pretreated in a 10-L steam-pretreatment unit, as described by Palmqvist et al. [40]. The pretreatment conditions and the impregnation procedure, utilizing an aqueous solution of 0.2% H_2SO_4, are described in more detail by Linde et al. [32]. Steam pretreatment was carried out for 10 min at a temperature of 190°C. The process configuration for the 2G stand-alone ethanol plant is shown in Figure 10.

It was assumed that the reactor is operated with 20-bar steam, corresponding to a reactor temperature of 190°C, and is modeled as a continuous reactor with 10% higher steam consumption than an adiabatic unit to account for heat losses. Pretreated material is assumed to be cooled by flashing at 4 and 1 bar. The flashed steam contains volatile compounds and, after condensation and cooling to 45°C, is fed to the WWT unit. Part of the flashed steam at 4 and 1 bar is not condensed, but instead recirculated and mixed with the incoming wheat straw to preheat the straw to 96°C. After flashing, the pretreated material (slurry) is cooled from 100°C to 55°C before mixing with fresh water prior to SSF.

SSF and yeast cultivation in 2G

Yeast is assumed to be produced on-site, since SSF makes yeast recirculation difficult. As a consequence, part of the pretreated slurry is pressed and the liquid collected for use in yeast cultivation. The liquid is supplemented with molasses to achieve a sufficiently high sugar concentration before being used for cultivation. The conversion of sugar to yeast was assumed to be 0.5 g biomass/g fermentable sugar in the simulations, and 2 g/L yeast is added in SSF. The concentration of the incoming material to 2G SSF is set to 10% WIS, and the temperature to 35°C. Enzymes are presumed to be purchased and added to achieve an activity corresponding to 10 FPU/g WIS. The ethanol yield for the 2G fermentation process is set to 0.84, with a sugar yield of 0.91 after enzymatic hydrolysis, and a glucose-to-ethanol conversion of 0.92, based on results reported by Linde et al. [32]. The ethanol concentration after the SSF in the broth that will be transfered to the distillation step would then be approximatly 2.7%.

Distillation in 2G

The 2G distillation process consists of two parallel 25-stage stripper columns operated at 3 and 1 bar (Murphree
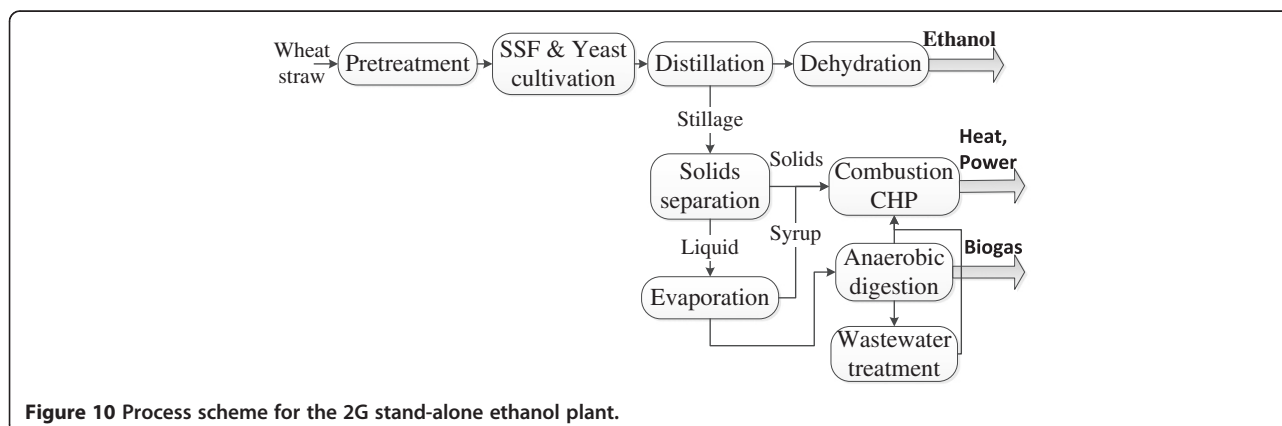

Figure 10 Process scheme for the 2G stand-alone ethanol plant.

efficiency 50%), and a 45-stage rectifying column operated at 0.25 bar (Murphree efficiency 75%). The SSF broth is divided into two equal streams and preheated to feed temperatures of 123°C and 90°C, before being sent to the stripper columns. After stripping, the overhead vapors are condensed and subsequently mixed with the rejected stream from the molecular sieve before being transferred to the rectifier. The intermediate stream fed to the rectifier would contain an ethanol concentration of about 26% and the reflux ratio for the non-heat integrated and the heat integrated case rectifier column was about 1.3 in both cases. The overhead vapor from the rectifier is heated from 50°C to 75°C and then fed to the molecular sieve, where it is assumed to reach an ethanol concentration of 99.5% before being condensed and cooled to 35°C. The rejected stream is condensed at 52°C before being recirculated to the rectifier. Thick stillage from the stripper columns is mixed and cooled to 80°C, as in the 1st generation plant, and the rectifier stillage is cooled to 45°C.

Evaporation in 2G

In the model a large part of the process water was assumed to be recirculated back to the simultaneous saccharification and fermentation step. This water normally contains compounds generated in the pretreatment step that can be toxic to the yeast if present in large amounts. Evaporation was therefore used to help reduce the toxic compounds present in the water, which otherwise will inhibit the yeast in the fermentation step. Potentially, syrup from the 2G evaporation step can also be diverted to the DDGS to increase the sugar amount in the product. Evaporation and the separation of solids are carried out in the same way as described for the 1G process. However, in this configuration, the syrup and the wet cake are burned in a CHP unit.

Anaerobic digestion and waste water treatment

It is assumed that biogas is produced from the condensed flash steam from the pretreatment step, from rectifier stillage, and from evaporation condensate, using anaerobic digestion. Since a limited amount of experimental data is available regarding this type of mixed substrate, there is some uncertainty in the modeling of the anaerobic digestion and WWT steps. It is assumed that the methane production corresponds to 0.35 m^3/kg consumed chemical oxygen demand [41]. The amount of consumed chemical oxygen demand in the biogas reactor is calculated assuming that easily digested material, such as organic acids and monomeric sugars, have a fractional conversion of 90%, compounds that require additional hydrolysis, such as cellulose and hemicellulose, have a fractional conversion of 50%, and that materials such as solid lignin, which are very difficult to hydrolyze are not converted.

Description of the combined 1st and 2nd generation plants

The 1G and 2G ethanol plants were integrated in two different ways. In the first case, only heat integration between the stand-alone plants, i.e. utilization of heat and sink sources, was considered. This is referred to as the co-located plants case. In the second scenario, both the energy and material streams were integrated by combining the fermentation broths prior to distillation, and altering the distillation step and the final products compared to the stand-alone cases. This is referred to as the combined plant case. In the latter case no DDGS are produced since the broths from the 1G and 2G plants are mixed prior to distillation. The thick stillage is sent for separation and evaporation before being incinerated, and thus, no dryer is needed. The process configuration for the combined 1G and 2G plant is shown in Figure 11.
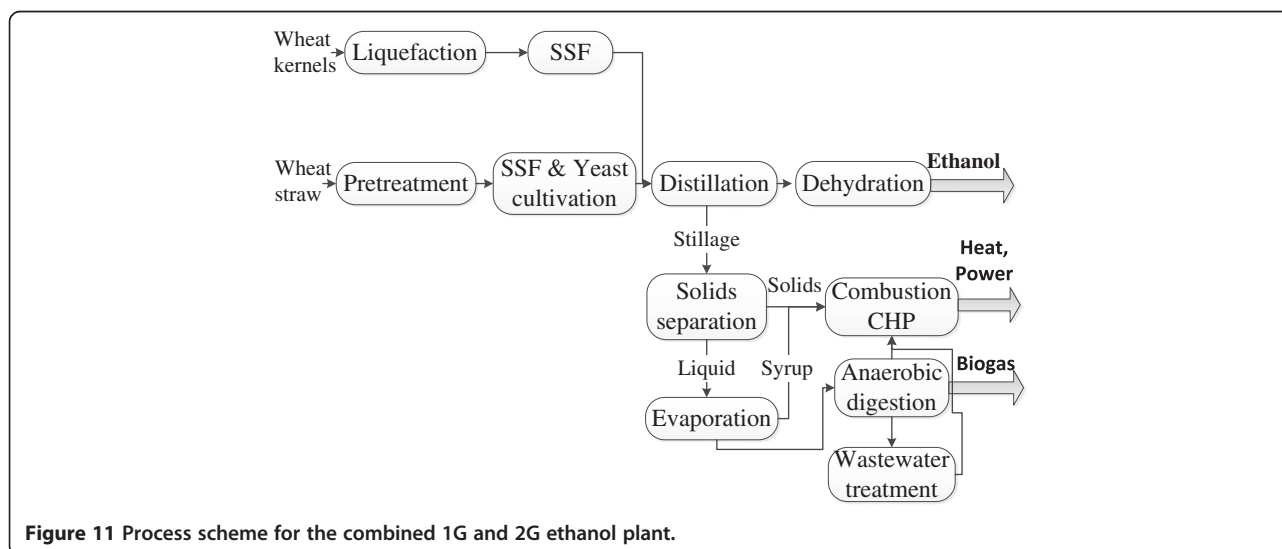

Figure 11 Process scheme for the combined 1G and 2G ethanol plant.

Table 6 The composition of wheat kernels and wheat straw (% dry basis)

	Wheat kernels [32,44-46]	Wheat straw [31]
Starch	66	0
Glucan	9	36
Xylan	7	20
Galactan	0	1
Arabinan	0	3
Protein	13	3
Fat	3	0
Acetyl	0	2
Extractives	0	3
Lignin	2	27
Ash	2	2

The distillation step is modeled using one stripper column with a top-stage pressure of 1.25 bar, and a rectifying column operating at 0.25 bar. The Murphree efficiency and the number of stages are the same as in the 2G scenario. The SSF broth from the combined plant is heated from 35°C to 100°C before entering the stripper column. Overhead vapor from the stripper is condensed and mixed with the reject stream from the dehydration step before being fed to the rectifier. The ethanol concentration in the stream fed to the rectifier was about 42% and the reflux ratio in the rectifier column was about 1.2 in both the non-heat integrated and the heat integrated case. After distillation, the vapor is heated to 72°C, dehydrated, and cooled to 35°C.

Energy supply

Steam at three different pressures is used in the simulations of utilities, depending on the configuration. For the 1G stand-alone plant, a moderate pressure of 6 bar and

Table 7 Stream data for the 1G stand-alone ethanol plant

Stream no.		Flow (kg/s)	T_{in} (°C)	T_{out} (°C)	Heat load (MW)
	Hot streams	27	90	37	4.6
1	Liquefied stream to SSF				
2	Stillage from stripper to separation	27	105	80	2.6
3	Condensing overhead vapor stripper column	6	92	81	8.9
4	Stillage from rectifier to WWT	3	142	45	1.1
5	Condensing 99.5% ethanol stream	3	130	114	2.3
5	99.5% ethanol to storage	3	114	35	0.6
6	Condensate from evaporation to WWT	15	81	45	2.1
7	Condensing overhead vapor from last evaporation step	5	63	60	11.5
7	Condensed vapor from last evaporation step to WWT	5	60	45	0.3
8	Condensing stream from molecular sieve	1	130	113	1.2
9	Condenser 1G rectifying column	4	113	112	13
	Cold streams	35	30	32	0.2
1	Temperature adjustment of liquefied stream to SSF				
10	SSF broth to stripper column	33	32	91	7.1
11	Fresh water to liquefaction	4	10	45	0.6
3	Condensate from stripper to rectifier	6	81	119	0.8
12	Overhead vapor rectifier to dehydration	4	113	130	0.1
13	Thin stillage after separation to evaporation	22	80	128	4.4
13	Evaporation of thin stillage: first step	22	127	128	6.6
14	Fresh air preheating to air mixing	35	10	20	0.4
15	Air mixture to dryer	124	70	150	10.7
18	Preheating of wheat kernel and water for liquefaction	26	38	60	1.8
19	Live steam heating in liquefaction	26	60	94	2.8
20	Reboiler stripper column	27	104	105	10.1
21	Reboiler rectifying column	3	141	142	16.6

a low pressure of 4 bar are assumed to be required; 6 bar is chosen since the dryer is designed to be operating at 150°C. In the 2G plant, high-pressure steam at 20 bar and low-pressure steam at 4 bar are chosen since pretreatment is carried out at 190°C, and the rest of the process is operated below 140°C. In the combined plant it is presumed that steam at all three pressures is available from the beginning of the process. It is assumed that cooling water between 10-19°C at a pressure of 1 bar is used in all plants. In all cases except the 1G stand-alone plant, it is also assumed that a CHP plant supplies the ethanol plants with steam. Primary steam is designed to be produced at 91 bar and 470°C, and electricity is produced by a steam turbine and a back-pressure turbine. The CHP plant is not included in the pinch analysis. The CHP system is designed to be used to incinerate residual material

(excluding any DDGS production) and biogas produced in the process. The residual material consists of a mixture of solids from the separation step after distillation and syrup from evaporation. When the incineration of residual material and biogas is not sufficient to run the process, natural gas, assumed to be pure methane, was added. The 1G stand-alone plant is presumed to be entirely powered by natural gas, as this is relatively common in the corn ethanol plants operating in the USA today [42]. Optionally, straw could have been used as external energy, however, methane was used in this case for easy comparison with existing infrastructure. Other sources of energy can be used to run the 1G plant. For example, sugarcane ethanol plants are often supplied with energy from the incineration of bagasse [43] and residual materials from the field, while at the 1G wheat ethanol plant in

Table 8 Stream data for the 2G stand-alone ethanol plant

Stream no.		Flow (kg/s)	T_{in} (°C)	T_{out} (°C)	Heat load (MW)
	Hot streams	0.4	190	80	0.64
22	Flash from pretreatment to WWT mixed stream				
23	Condensing 4-bar flash steam from pretreatment	1.3	144	140	2.5
23	Cooling 4-bar flash steam to WWT mixed stream	0.2	99	93	0.34
24	Slurry to SSF	15.4	100	55	2.08
25	Liquid fraction to yeast cultivation	2.2	35	30	0.04
26	Stillage from low-pressure stripper to separation	17.7	110	80	2.13
27	Stillage from high-pressure stripper to separation	19.6	136	80	4.53
28	Condensing overhead vapor: low-pressure stripper	2.3	103	92	4.56
29	Condensing overhead vapor: high-pressure stripper	2.6	130	117	4.97
30	Condensing rejected stream molecular sieve	0.4	75	52	0.53
31	Condensing 99.5% ethanol stream	1.1	75	50	1.04
31	99.5% ethanol to storage	1.1	50	35	0.05
32	Condensate from evaporation to WWT	21.1	81	45	2.82
33	Condensing overhead vapor from last evaporation step	6.9	63	60	16.05
33	Condensed vapor from last evaporation step to WWT	6.9	60	45	0.42
34	WWT mixed stream from pretreatment to WWT	1.8	74	45	0.34
35	Stillage from rectifier to WWT	3.8	86	45	0.65
36	Condenser 2G rectifying column	1.5	50	49	2.04
	Cold streams	20.1	35	103	5.2
37	SSF broth to low-pressure stripper column				
38	SSF broth to high-pressure stripper column	22.3	35	130	8.3
36	Overhead vapor rectifier to dehydration	1.5	50	75	0.1
39	Thin stillage after separation to evaporation	31.9	80	129	6.7
39	Evaporation of thin stillage: first step	31.9	128	129	8.4
40	Live steam for wheat straw in reactor	15.4	97	190	6.3
41	Reboiler low-pressure stripper column	17.7	109	110	5
42	Reboiler high-pressure stripper column	19.6	135	136	5.3
43	Reboiler 2G rectifying column	3.8	85	86	3.1

Norrköping, Sweden, biomass and domestic waste are used to produce steam [44]. The electricity needed in the plants was included when calculating the energy need to produce one kg of ethanol, by assuming an electricity requirement of 0.84 MJ/kg ethanol. The assumption was made as there are limited data available concerning the energy used for electricity in the considered cases. This value corresponds to an average electricity demand described in a 1G ethanol corn plant [29].

Raw material

For the composition of wheat kernels, average values were calculated based on data compiled by Erdei et al. [33] and data found in the literature [45-47]. Data on the composition of wheat straw were taken from a study performed by Linde et al. [32] for straw collected in

southern Sweden. The wheat straw and kernels were assumed to have dry matter contents of 88% and 86%, respectively. The dry matter compositions of straw and kernels are presented in Table 6.

Base case

Four base cases were analyzed: a stand-alone 1G plant and a stand-alone 2G plant, a co-location of a 1G and a 2G stand-alone plant in which only energy was considered for integration, and finally a combined plant in which both the energy and material streams of a 1G and 2G plant were combined. In the combined case SSF broths from the 1G and 2G plants were mixed prior to distillation.

It was assumed in the base case that the streams in the plants were heated and cooled with utilities only.

Table 9 Stream data for the combined 1G and 2G ethanol plant

Stream no.		Flow (kg/s)	T_{in} (°C)	T_{out} (°C)	Heat load (MW)
	Hot streams	27.8	90	38	4.5
44	Liquefied stream to SSF				
45	Flash from pretreatment to WWT mixed stream	0.4	190	80	0.6
46	Condensing 4-bar flash steam from pretreatment	1	144	140	2
46	Cooling 4-bar flashed steam to WWT mixed stream	0.2	99	94	0.3
47	Slurry to SSF	15.3	100	55	2
48	Hydrolysate to yeast cultivation	2.2	35	30	0
49	Stillage from stripper to separation	64.6	110	80	7.3
50	Condensing overhead vapor from stripper	10.7	100	89	18.4
51	Condensing rejected stream molecular sieve	1.4	72	48	1.9
52	Condensing 99.5% ethanol stream	3.9	72	47	3.7
52	99.5% ethanol to storage	3.9	47	35	0.1
53	Condensate from evaporation to WWT	37.2	81	45	5.2
54	Condensing overhead vapor from last evaporation step	12.1	63	60	28.4
54	Condensed vapor from last evaporation step to WWT	12.1	60	45	0.7
55	WWT mixed stream from pretreatment to WWT	1.5	90	45	0.3
56	Stillage from rectifier to WWT	6.8	84	45	1
57	Condenser rectifying column	5.3	47	46	6.6
	Cold streams	4.5	10	45	0.6
58	Fresh water to liquefaction				
59	Temperature adjustment of liquefied stream to SSF	35.7	31	32	0.1
60	SSF broth to stripper column	75.4	34	100	18.2
61	Overhead vapor rectifier to dehydration	5.3	47	72	0.2
62	Thin stillage after separation to evaporation	53.9	80	129	10.7
62	Evaporation of thin stillage: first step	53.9	129	133	16.7
64	Live steam for wheat straw in reactor	15.4	97	190	5.8
65	Live steam heating in liquefaction	26.6	60	94	2.9
66	Preheating of wheat kernels and water for liquefaction	26.6	38	60	1.8
67	Reboiler stripper column	64.6	109	110	20.5
68	Reboiler rectifying column	6.8	83	84	11.4

The base cases were compared with cases in which utilities and energy integration, through heat exchange, supplied the process with the energy necessary for heating and cooling. This resulted in eight different cases; the four base cases, and the four cases with utilities and energy integration by heat exchange.

Pinch analysis

Aspen Energy Analyzer (AEA) version 7.2 was used to design heat exchanger networks, locate the pinch temperature and calculate the energy targets for each of the simulated scenarios. The program can produce recommended network designs and perform cost estimates automatically. However, these features were not utilized in this study since the automatically generated network designs do not always consider pinch violation, stream splitting, the location of streams and the condensation and evaporation of streams in a satisfactory way. The automatic cost estimates were not used since they are based on older data and not sufficiently flexible to changes. Instead, Aspen Process Economic Analyzer (APEA) version 7.2.1 was used to estimate the investment cost of heat exchanging equipment.

The initial and final temperatures, the flow rates and the heat loads of the streams were exported from Aspen Plus. The overall mass heat capacity was calculated using the heat load and the temperature before being inserted into AEA version 7.2. The global minimum temperature difference (ΔT) in the heat exchangers was set to 10°C, but lower and higher temperature variations were applied to specific streams. No changes, such as varying the pressures in the distillation or evaporation stages, or modifying the drying gas in the dryer, were made to the original processes. These investigations can be performed by extracting the selected stages from the analysis, the remaining energy flows will then be known as the background process.

Some restrictions and considerations were taken into account when modeling the heat exchanger network to make the simulations more applicable in real industrial processes. Heat exchange was not allowed between evaporating or condensing streams unless the stream in question was completely evaporated or condensed during heat exchange. Stream splitting was kept to a minimum by only allowing a stream to be split once. The physical locations of the streams in the facility must be considered when modeling the heat exchanger networks. In this study, heat exchange in the CHP plant or between the ethanol and CHP plant was omitted from the evaluation, also no heat exchange was carried out between the effects inside the evaporation stage and the rest of the process; in this case only the incoming and outgoing streams from the unit were included in the network. No heat exchange was allowed in stages where live steam was used, i.e. pretreatment and liquefaction;

however, the heat required was still included since this indicates the potential for live steam replacement with secondary steam. The streams considered in the pinch analysis for the different scenarios are listed in Tables 7, 8 and 9. The co-located case is simply a combination of the 1G and 2G stand-alone cases, and is therefore not described separately.

Cost calculations

Estimates of the cost of heat exchangers were obtained using APEA version 7.2.1, using the heat transfer areas generated in AEA version 7.2, and temperature settings from Aspen Plus. The maximum design pressure was set to 10 bar. The tube and shell material were chosen to be stainless steel, SS316. The heat exchanger area was overestimated by 15%. Thereafter, the total direct costs (equipment and installation cost) were calculated for all eight scenarios assuming an European plant location. The total direct costs were discounted using the annuity method with an internal rate of return of 7% and an economic lifetime of 15 years.

The cost of utilities was calculated based on the total amount of cooling water, 4-, 6- and 20-bar steam calculated in the Aspen Plus simulations, and the following energy prices: cooling water 0.016 EUR/ton, 4- and 6-bar steam 20 EUR/ton, and 20-bar steam 25 EUR/ton.

The discounted total direct cost was added to the total annual cost of utilities to give an annual total cost. The total investment cost for the plant was not calculated, and no extra costs for piping were included.

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