

Effects of process decarbonisation on future targets for excess heat delivery from an industrial process plant

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Abstract

The use of industrial excess heat for purposes such as district heating has the potential to contribute to societal targets for energy efficiency and greenhouse gas emissions reduction. However, to meet the ambitious national and international climate targets set for 2050, a breadth of different decarbonisation pathways are required, not least in the industrial sector. These include a transition to bio-based and recycled feedstock and fuels, carbon capture and storage, and electrification. Such profound changes of industrial processes and energy systems are likely to affect the availability of excess heat from these plants, and a better understanding of how the excess heat potentials might change is needed in order to utilise excess heat in ways that can be resource-efficient also in the long-term. In this paper, we present a systematic approach which can be used to analyse how different decarbonisation options may affect the potential future availability of excess heat at a specific plant site. The approach is based on the use of consistent, energy targeting methods based on pinch analysis tools, and therefore relies on comprehensive data about process heating and cooling demands. To illustrate the approach, we demonstrate results from two industrial case studies in which different decarbonisation measures are assumed to be implemented. The case studies were selected from a case study portfolio, which includes relevant and site-specific process and energy data for a large share of Swedish industrial process sites. The results show that deep decarbonisation can have significant impact on the availability and temperature profile of industrial excess

heat, illustrating the importance of accounting for future process development when estimating excess heat potentials.

Introduction

The use of industrial excess heat (e.g. for district heating) is recognised in the EU Energy Efficiency Directive (European Parliament and Council of the European Union 2012) as an important resource for reaching energy and climate targets, and several studies (Forman *et al.* 2016; e.g. Panayiotou *et al.* 2017) indicate a large potential for increased utilisation of this resource. Meanwhile, several roadmaps and pathways for decarbonisation of industry have recently been presented (e.g. for European industry (de Vita *et al.* 2018; Fleiter *et al.* 2019), for the global industry (de Pee *et al.* 2018), for eight industrial sectors in the UK (WSP Parson Brinkerhoff and DNV GL 2015), for the chemicals sector of the Netherlands (Stork *et al.* 2018), and for nine industrial sectors in Sweden (Sweco 2019)). These roadmaps highlight priority areas such as electrification, increased use of renewable and recycled raw materials and carbon capture and storage (CCS), all of which would significantly affect the availability of excess heat from industrial processes.

However, assessments of industrial excess heat potentials are generally based on current energy balances of existing industrial plants, for example by estimating excess heat availability as an experience-based, sector-specific fraction of the energy supplied by combustion (e.g. Papapetrou *et al.* 2018; Persson *et al.* 2014). By not accounting for future process changes, such estimates risk leading to the wrong conclusions and decisions about the best future use of industrial excess heat. A better understanding of how the excess heat potentials might change

is needed in order to utilise excess heat in a way that can be resource-efficient also in the long-term, and that avoids future lock-in effects.

For assessments of plant-specific excess heat potentials, it is appropriate to use methods based on pinch analysis, which explicitly account for process heat recovery opportunities by considering temperatures and heat loads of individual process heat sources and sinks (see also Theoretical background). Recently, Klemeš *et al.* (2018, Section 2.6.2.2) reviewed the development and application of pinch-based methods to excess heat recovery assessments. More recently, Svensson *et al.* (2019b) presented a systematic method for characterisation and visualisation of industrial excess heat that allows for consistent and clear assumptions about the trade-offs between on-site integration of combined heat and power units, and excess heat recovery. Additionally, Pettersson *et al.* (2020) presented a holistic methodology for assessing the cost- and environmental benefits of industrial excess heat utilisation in different time perspectives. In parallel with the development and applications of excess heat assessment methodologies, numerous studies have investigated the integration of new technology and processes that may contribute to greenhouse gas emissions reductions in existing industrial plants. A few examples from ongoing or recently completed projects include industrial carbon capture (Andersson *et al.* 2016), industrial electrification (Wiertzema *et al.* 2018), and the integration of new biobased processes (Ahlström *et al.* 2017). Together, these research projects provide an excellent source of knowledge, methods and tools, as well as data for analysing future potentials for excess heat from decarbonised industrial processes. However, to be able to provide a comprehensive picture of how the availability of industrial excess heat may change in a future with close to net-zero emissions from industry, a systematic approach that allows for comparing and aggregating the results from different case studies is needed.

The aim of this paper is to present a systematic methodology based on consistent assumptions for analysing – at the plant level – how technological development of industrial process plants can affect the potential availability of excess heat. The methodology is demonstrated through two industrial case studies in which technological changes for deep emissions reductions are assumed to be implemented at the studied plants. The suggested methodology is a first step towards a bottom-up approach for estimation of future potentials for industrial excess heat.

Theoretical background

Heat recovered from an industrial process site can be valorised in different ways to be utilized for new purposes, e.g., as a heat source for a new process or district heating network, or for low-temperature power production. Thereby it can replace other heat sources and reduce the demand for primary heat supply. Such utilisation of excess heat competes with opportunities for enhanced on-site process heat recovery to reduce the demand for primary heating and cooling to the industrial processes. On the other hand, it is usually not possible to recover all heat within an industrial process site via direct heat exchange due to the thermodynamic constraints set by temperatures of heat sources and sinks. This means that even after maximised theoretical heat recovery, there is still – unavoidable – excess heat

available from most processes (Bendig *et al.* 2013). This heat can be valorised internally, e.g. through heat pumping or low-temperature power production or recovered for off-site deliveries without further limiting the possibilities to save fuel on-site. The use of avoidable excess heat can affect the overall system energy efficiency, fuel use and greenhouse gas emissions either positively or negatively depending on the trade-off between on-site and off-site fuel savings (Eriksson *et al.* 2018). The use of unavoidable excess heat, on the other hand, inevitably implies an improvement in overall system performance.

Excess heat at the plant-level is commonly estimated using methods and tools based on pinch analysis (Klemeš 2013; Smith 2016). These methods account for thermodynamic as well as techno-economic constraints and provides ways to analyse and visualise the potential for improvements in heat recovery within and between processes at the industrial site. The application of pinch analysis requires temperature-specific data for process heating and cooling demands, and an assumed value for the minimum acceptable temperature difference for heat exchange, ΔT_{\min} . These are then used to calculate the optimal heat cascade of the system to maximise process heat recovery and determine the corresponding minimum requirements for external heating and cooling of the process. The ideal heat cascade is visualised in the Grand Composite Curve (GCC), which is one of the main graphical tools of pinch analysis. The GCC can be used for visualising the potential for heat integration between a process and its utility system or another process. The unavoidable excess heat from a process corresponds to its minimum theoretical cooling demand, which can be read directly from the GCC for a given value of ΔT_{\min} . A reference condition representing the very theoretical maximum potential for process heat recovery is given by assuming a ΔT_{\min} of 0 °C.

In many common energy-intensive manufacturing processes, heat is available at very high temperatures from process operations. For example, there can be a demand for cooling of process streams at high temperatures resulting from thermochemical conversions or chemical reactions. Other processes involve the generation of various energy-rich by-products that are difficult to transport, store and sell, and are therefore most conveniently combusted on-site, resulting in high-temperature heat generation that, in principle, can be seen as unavoidable. Examples include the light vapours used as fuel gas in oil refineries, the blast furnace gas and other off-gases from iron and steel manufacturing processes, and the black liquor from Kraft pulping industry. Steam boilers and furnaces fuelled with by-products that would otherwise be flared or destructed, high-temperature heat generation in reactors, and combustion processes that are inherently related to process unit operations (e.g. lime kilns in the cement industry or pulp industry) can be seen as unavoidable heat sources.

Furthermore, most industrial process sites have a central utility system, which supplies heating and cooling to the process. This utility system is very commonly a steam system, possibly with integrated steam turbine cycles for co-generation of shaft power or electric power. In addition to excess heat from the process itself, any residual heat from such an integrated utility system should also be recognised (see e.g. Oluleye *et al.* 2016).

In conclusion, the assessment of excess heat will depend on the availability of high-temperature heat and the steam system

configuration, which in turn depend on site-specific priorities between fuel savings, co-generation of shaft/electric power and total (on-site) system efficiency. It also becomes necessary to estimate, not only targets for minimised heating and cooling demands, maximised heat recovery and potential excess heat availability, but also the power production target.

Methodology

In this article we present a systematic methodology to estimate and analyse excess heat availability under current and potential future process development scenarios. The methodology follows the steps outlined below.

STEP 1. DATA COLLECTION

This step involves the collection of data about heat loads and temperatures of process heat sources and sinks, including potential high-temperature heat sources such as heat from combustion of off-gases or non-marketable by-products or exothermic reaction heat.

In this paper, the methodology is illustrated by two case studies, for which the necessary data for the existing process sites were taken from the Chalmers Industrial Case Study Portfolio (ChICaSP). ChICaSP is a collection of case study data that has recently been compiled based on previous research projects at Chalmers (Svensson *et al.* 2019a). It includes relevant and site-specific process and energy data for a large share of Swedish industrial process sites. However, the high-temperature heat from reactions or internally generated fuel by-products was not always considered in previous case studies (or was regarded as a utility) and required some additional data assumptions or estimations in this work. Data for new processes were based on literature.

STEP 2. IDENTIFICATION OF OPTIMAL HEAT CASCADE FOR EXISTING AND DECARBONISED PROCESS

As discussed in the Theoretical Background, the assessment of optimal heat recovery in systems including high-temperature heat-sources and co-generation requires a prioritisation between fuel consumption, co-generation and total system efficiency. In the present work, we followed the prioritisation order below, which is based mainly on the prioritisation proposed in (Svensson *et al.* 2019b):

1. minimise heat production in utility boilers and furnaces (fired by purchased fuel),
2. maximise co-generation in back-pressure steam turbines given process steam demands at existing steam levels and the minimised heat production,
3. maximise condensing power production provided that this does not reduce the excess heat available for covering the current capacity for excess heat deliveries,
4. maximise excess heat recovery for new purposes (e.g. district heating, heat pumping or low-temperature power production).

The optimal heat and power integration can be visualised using common pinch-based tools. The minimisation of hot utility (priority 1) can be illustrated in a process GCC, which include

process-inherent high-temperature heat sources. The maximisation of steam turbine power generation (priorities 2 and 3) can be visualised by split GCCs, where the steam turbine cycles are represented as a foreground GCC curve, which is matched against the process GCC in the background. Finally, the excess heat (priority 4) can be read from a GCC that represents a combination of the process and steam turbine cycles. In this work, we made some further simplifications to the representation of the excess heat temperature profile (see Step 3).

In addition to the priorities listed above, a ΔT_{\min} of 0 °C was assumed in this work. Altogether, these assumptions represent a highly theoretical reference case for heat recovery.

In practice, operability, safety, and plant layout put constraints on potential process integration. However, for new, emerging processes, consideration of such constraints would require extensive additional assumptions about how the plant will be constructed and how that affects the extent of heat integration. To avoid speculations and enable consistent assumptions, the theoretical case of ideal heat recovery, and priorities according to the list above, were applied in this study. This case represents reference conditions for the absolute limit for further energy efficiency improvements through direct heat recovery and co-generation and, consequently, serves as a conservative estimate of excess heat availability. Note, however, that possibilities for heat pumping for increased internal heat recovery are not explicitly represented in the trade-off targeting approach. Instead, heat pumping opportunities are considered as one of several excess heat utilization options.

The energy targeting computations were performed using Mat4PI (Morandin 2017), an in-house program in which the targeting problem is solved as a linear programming problem where primary hot utility, power generation and excess heat are given different weights in the objective function to reflect the prioritisation above, i.e. minimised fuel use is given highest priority, and maximised power generation is given second priority. Excess heat is prioritised last and furthermore weighted according to its temperature level (see also Step 3). For a more detailed description and discussion about the weighting and other targeting assumptions, see Svensson *et al.* (2019b).

STEP 3. COMPARISON OF RESULTING HEAT LOAD-TEMPERATURE PROFILES FOR UNAVOIDABLE EXCESS HEAT

To visualise excess heat availability and temperature characteristics, a newly developed graphical tool called the excess heat temperature (XHT) signature was used (Svensson *et al.* 2019b). The XHT signature illustrates the amount of excess heat available according to pre-defined temperature ranges and constant temperature levels that represent relevant excess heat utilization options such as district heating, heat pumping or low-temperature power production cycles. It is an appropriate tool for comparing excess heat availability from different processes (e.g., as in this study, current and future process alternatives) since an XHT signature can be defined and constructed in a systematic way given consistent assumptions. Compared to the GCC, the XHT provides a more direct interpretation of the excess heat availability, which facilitates comparison of excess heat potentials from different sites. Furthermore, multiple XHT signatures can easily be aggregated (by addition) into one curve to illustrate the total excess heat availability for a group of plants.

Table 1. Temperature ranges and weights used for construction of the XHT signatures.

Temperature range	Weight factor	Temperature range	Weight factor	Temperature range	Weight factor
250 °C–250 °C	3.62	110 °C–120 °C	2.24	85 °C–85 °C	1.80
135 °C–250 °C	3.06	95 °C–110 °C	2.07	60 °C–85 °C	1.58
135 °C–135 °C	2.51	95 °C–95 °C	1.96	60 °C–60 °C	1.36
120 °C–135 °C	2.41	85 °C–95 °C	1.88	25 °C–60 °C	1.00

The assumptions for process heat recovery and co-generation described in Step 2 above, correspond to reference conditions for a Theoretical XHT signature as suggested in (Svensson *et al.* 2019b). This Theoretical XHT signature thus represents the maximum amount of excess heat that can be made available without limiting the potential for further primary heat savings or steam turbine power generation on-site and therefore represents the unavoidable excess heat.

In this study, we used the temperature levels and ranges specified in Table 1 to define the XHT signatures. The weight factors were calculated based on Carnot efficiencies at the average temperature within each temperature range.

Case Study 1: Steam cracking

This case study focused on the implications of replacing the steam cracking plant currently used for production of light olefins (mainly ethylene and propylene) at a large petrochemical cluster in Sweden, with new processes utilising lignocellulosic biomass feedstock (hereafter referred to as BTO – biomass-to-olefins).

EXISTING PROCESS DESCRIPTION: STEAM CRACKER

The studied steam cracker site is the heart of the largest petrochemical cluster in Sweden. Total olefin production exceeds 700 ktonnes/year with an ethylene-to-propylene mass ratio (E/P-ratio) of 3.25 (Table 2). The steam cracker site consists of a cracking section followed by recovery of the main products, ethylene and propylene. By-products from the recovery section are upgraded to steam cracked naphtha (SCN) and ethylene-tert-butyl-ether (ETBE).

In the cracking section, light hydrocarbon feedstock is mixed with steam and heated to 850 °C in nine cracking furnaces, in which the feedstock is cracked into unsaturated molecules, mainly ethylene. The raw product gas is cooled and sent to ethylene recovery, where several side streams are obtained in addition to ethylene. C3 hydrocarbons are reacted to propylene while C4 hydrocarbons are used for ethylene-tert-butyl-ether (ETBE) production. Steam cracked naphtha (SCN) and fuel oil are recovered from C5+ hydrocarbons.

Steam for process heating is generated by cooling of product gas from the cracker furnaces and by local utility boilers. Steam at 85 bar(g) is expanded in back-pressure turbines for generation of electric power and shaft work to process equipment, before being used for process heating at 8.8 or 1.8 bar(g). Currently, about 30 MW of shaft work is generated in direct drive turbines, while about 10 MW of electric power is generated in a turbo-alternator. Steam at 40 bar(g) is exported while steam at 1.8 bar(g) is imported (about 5.2 and 29.1 MW, respectively (Johansson 2013)).

The utility boilers and the cracker furnace are fired by fuel gas – mainly hydrogen and methane – generated in the process. Natural gas is used as make-up fuel if necessary and excess fuel gas is exported to the other plants in the cluster. In this paper, it was assumed that all fuel gas must be used within the cluster, as there are currently no options for export outside the cluster.

Thermal stream data for the steam cracker was primarily based on a pinch analysis study conducted by Hedström and Johansson (2008). The cracker furnaces were not included in their analysis and streams representing this part of the process were instead taken from ChICaSP (Svensson *et al.* 2019a) and are consistent with the data used by Hackl *et al.* (2010). Other changes to the stream data given by Hedström and Johansson (2008) included a change in the heat duty of the quench oil stream (using updated stream data from (Svensson *et al.* 2019a)), and replacement of streams representing feedwater heating with streams representing generation of direct injected process steam. Generation of exported steam was included in the cold stream data, while imported steam was included as a hot stream at its condensation temperature. Furthermore, the internal fuel system in the cluster is in balance meaning fuel exports to the cluster cannot be increased. Consequently, the current fuel gas consumption was considered unavoidable and was included in the stream data.

NEW PROCESS DESCRIPTION: BIOBASED OLEFINS

The bio-based olefins production considered in this case study is a combination of the methanol-to-olefins (MTO) and the ethanol-to-ethylene (ETE) processes. The required alcohols – methanol and ethanol – are produced on-site from biomass feedstock.

The assumed process for production of biobased olefins via methanol was based on the process described in Arvidsson *et al.* (2016), while the process assumed for ethylene production via ethanol was based on the process described in Arvidsson and Lundin (2011). Note that the MTO process produces both target olefins – ethylene and propylene – with mass ratios ranging from 0.7 to 1.4 depending on employed operating conditions. This is far lower than the E/P-ratio of the current cracker product. The ETE process is used to shift the overall production towards ethylene. The relative sizing of the two processes depends on the E/P-ratio of the MTO-process, with a lower E/P-ratio requiring a larger ETE process. In the present work, a low E/P-ratio of 0.76 was used since a number of studies indicate that it could be cheaper to produce ethylene via the ETE-process (see e.g. Bazzanella & Ausfelder 2017; Johansson & Pettersson 2014).

The studied process for bio-based olefins production via the MTO-process starts by drying and torrefaction of biomass, followed by oxygen-blown entrained flow gasification. The

resulting syngas is conditioned and used for methanol synthesis. In the subsequent MTO-process, methanol and steam are processed in a catalytic circulating bed reactor at roughly 400 °C and 4 bar(a) to produce ethylene and propylene. The catalytic bed material is continuously regenerated by burning of coke formed in during the reaction. Ethylene and propylene are recovered in a separation section similar to that of conventional steam cracker plants. Various other combustible gases – including torrefaction gases – are recovered in the process. In the present work, it was assumed that the combustible gases will primarily be used to replace current fuel gas exports to the cluster, and that any remaining gases must be combusted on-site. A simplified process flow diagram is shown in Figure 1.

Thermal stream data for the methanol and MTO-processes was based on Arvidsson *et al.* (2016) with the modification that heat released due to by-product combustion was reduced by 70 MW to reflect fuel gas exports to the rest of the cluster. This value was derived from the current fuel gas exports of 82 MW and an assumed boiler efficiency of 85 %.

The studied process for ethylene production via lignocellulosic ethanol production is described in Arvidsson and Lundin (2011) and uses two-step dilute sulphuric acid steam explosion pre-treatment followed by enzymatic hydrolysis and co-fermentation (SSCF). The fermentation product is purified to 93 wt% ethanol in a two-step process consisting of a Beer column followed by a rectifier for water-ethanol separation. The Beer column bottom product consists of a lignin-rich solid fraction and a liquid fraction containing solubilised non-volatile compounds. The liquid fraction is concentrated in a five-eff

fect evaporation unit and mixed with the lignin-rich solid fraction of the Beer column. In the present work, it was assumed that the residue after mixing is pelletised and sold as fuel.

Produced ethanol is dehydrated to ethylene by an endothermic reaction carried out in four adiabatic reactors in series operated at 11.4 bar(a). Each reactor feed is heated to 450 °C in a furnace and superheated steam at 11.4 bar(a) is injected with the feed to the first reactor to serve as a heat carrier. The final reactor outlet is quenched by spraying water, causing condensation of the steam. The dry product gas is then cleaned from CO₂ and upgraded to polymer grade ethylene using a distillation column followed by a stripper. Heavy ends from the distillation column is the only process by-product and in the present work, it was assumed that this must be combusted on-site. A simplified process flow diagram is shown in Figure 2.

Thermal stream data for ethylene production via ethanol using the ETE-process was based on Arvidsson and Lundin (2011) and adjusted for the purpose of the present work by adding streams to represent by-product combustion and generation of process steam for direct injection.

The assumption that the solid residue of the ethanol process is upgraded to pellets implies an additional heat demand for drying the biomass to 10 % moisture content. To include this, the heat load of a biomass dryer required to reach 10 % moisture content, which was already included in the stream data for the gasification plant given in Arvidsson *et al.* (2016), was increased proportionally to the amount of additional water evaporated.

The required production of ethylene and propylene by the MTO and ETE processes for a complete feedstock switch is

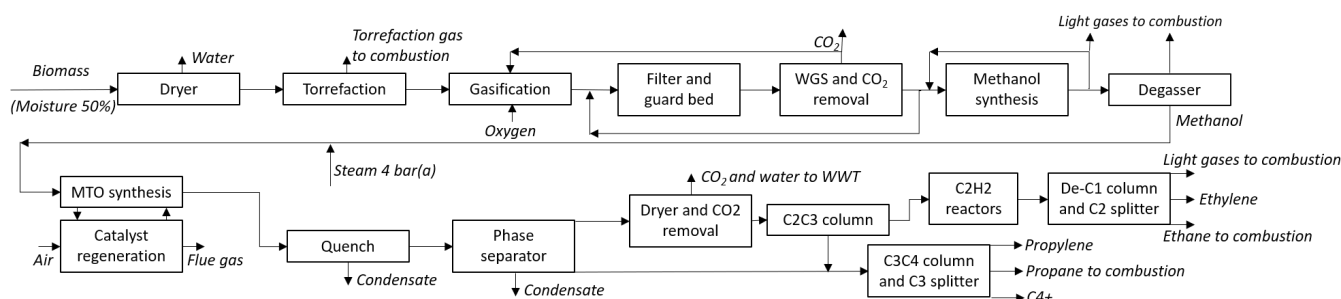


Figure 1. The process for bio-olefin production via the MTO-process, as modelled by Arvidsson *et al.* (2016).

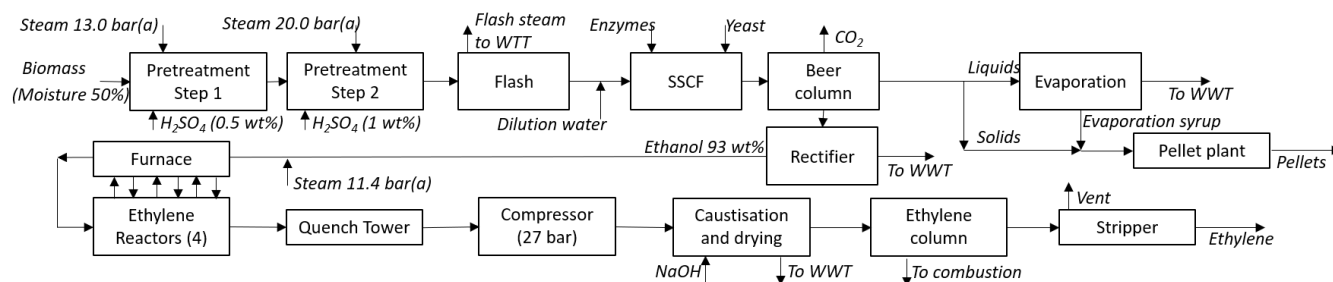


Figure 2. The process for bio-ethylene production via the ETE-process, as modelled by Lundin and Arvidsson (2011).

summarised in Table 2. The corresponding biomass and alcohol requirements are presented in the same table and have been estimated from the conversion efficiencies given in Arvidsson *et al.* (2016) for the MTO process and in Arvidsson and Lundin (2011) for the ETE process.

EXCESS HEAT EVALUATION

The GCC of the existing steam cracker plant is presented in Figure 3 (left), indicating that the external fuel demand is zero and that excess heat is available at high temperature. The integration between the CHP plant steam cycle and the steam cracker is also visualised in Figure 3 (left). There are currently no excess heat exports from the cracker (since steam exports are included in the stream data) meaning that steam extraction flows from the back-pressure turbines are determined by process heat demands, after which condensing power generation is prioritised. The power generation target is 65 MW, compared with the 40 MW of power currently generated at the site. The remaining excess heat can be visualised in the XHT signature of the site (Figure 4, solid line). The signature indicates 140 MW of excess heat, of which 120 MW is available in the low temperature interval (25–60 °C). The sharp step between 85 and 60 °C is explained by the presence of a heat pocket in the process GCC (barely visible at 200 MW in Figure 3 (left)).

The BTO-process has significant amounts of high-temperature heat available due to cooling of the gasification product gas stream. In addition, various fuel by-products are generated in the processes (see process description above), including a solid residue from ethanol production, a liquid fuel product from the ethylene process, and fuel gases from gasification, methanol production and the MTO-process. Even after fuel gas exports to the cluster and pellets production, large amounts of high temperature excess heat remain from the BTO-process. Since excess heat is not currently exported from the site, co-generation followed by condensing power production is prioritised after minimising fuel use. The resulting power generation target is 94 MW. The split GCC of the BTO-process and the integrated steam cycle is shown in Figure 3 (right). Remaining excess heat from the BTO-process is visualised by the process XHT-signature, shown in Figure 4 (dashed line) and indicating a total of just above 140 MW excess heat, mostly available at 95–85 °C, with about 30 MW at 110–95 °C and 6 MW at 60 °C.

Figure 4 compares the theoretical XHT signature of the BTO-process concept with that of the existing process. The total amount of excess heat is roughly unchanged in the BTO-process, but heat is available at considerably higher temperatures.

Current power production at the steam cracker is about 40 MW (of which about 10 MW is electric power and the remainder is shaft power for direct drive of process equipment). In the ideally integrated process with maintained on-site fuel consumption, the power generation target is 65 MW and in the BTO-process, it increases to 94 MW. The power consumption of the BTO-process was estimated to 223 MW for the BTO-process based on scaling of data provided in Arvidsson *et al.* (2016) and Arvidsson and Lundin (2011), indicating a theoretical deficit of 129 MW. In 2017, 342 GWh of electric power were imported to the existing cracker plant according to environmental reports. Assuming 93 % plant availability, this corresponds to a power deficit of 42 MW and a total power consumption of 82 MW. Comparing to the power generation target (65 MW), the theoretical deficit is only 17 MW.

Case study 2: Kraft pulp mill

In this case study we analysed the consequences of implementing carbon capture of the biogenic CO₂ emitted from a kraft pulp mill process. By transporting and permanently storing the CO₂, a net removal of CO₂ from the atmosphere is achieved, thus creating so-called negative emissions that may compensate for fossil emissions from other sectors.

PROCESS DESCRIPTION: SOFTWOOD KRAFT PULP MILL

The case study mill is a softwood Kraft pulp mill located on the Swedish west coast, with a production capacity of 700,000 tonnes of pulp per year. After recent expansion and retrofits, the mill is equipped with state-of-the-art technology, including a continuous digester, modern recovery boiler and a new condensing turbine. As part of the rebuild, a new secondary heating system for efficient heat recovery was also implemented.

The Kraft process is a chemical pulping process, by which the cellulose fibres are extracted from the wood using white liquor (a solution of sodium hydroxide and sodium sulphide). During cooking in the digester, the lignin content of the wood is dissolved and ends up in the black liquor stream, which also contains the spent cooking chemicals. The black liquor that is washed out from the pulp after cooking is sent to the evaporation plant, where water is evaporated to increase the dry solids content and allow the black liquor to be combusted in the recovery boiler. The main purpose of the recovery boiler is to regenerate cooking chemicals from the black liquor, but in the process, large amounts of energy are also recovered.

Table 2. Olefins production in new and existing processes, and feedstock consumption for new processes.

Process	Production (ktonnes/year)		Alcohol feed (ktonnes/yr)	Biomass feed ¹ (TWh/yr)	Hydrocarbon feed (ktonnes/yr)
	Ethylene	Propylene			
Existing steam cracker	565	174	–	–	1,183
MTO	133	174	1126	8.8	–
ETE	432	–	810	13.3	–
Sum (bio-processes)	565	174	1,936	22.1	–

¹ Assuming LHV=17.6 MJ/kg dry fuel.

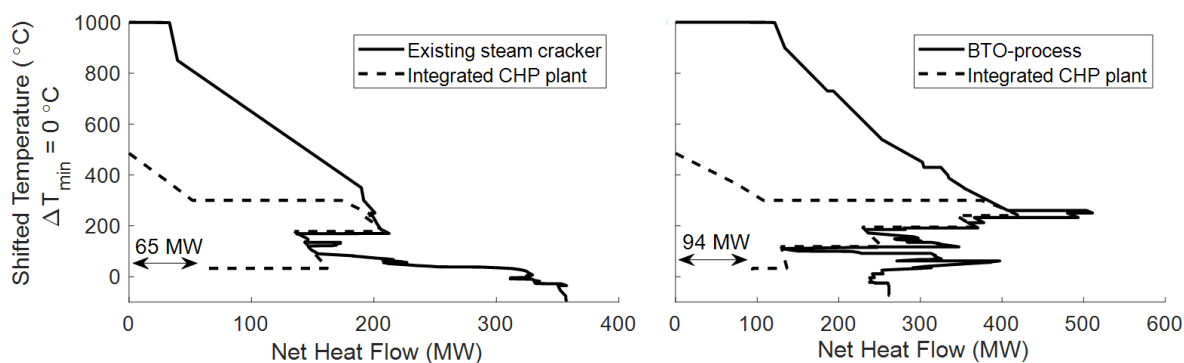


Figure 3. Left: Split GCC of the existing steam cracker plant and the steam cycle of the integrated CHP plant. Right: Split GCC of the BTO-process and the steam cycle of the integrated CHP plant.

Bark is a by-product from both the pulping process and the sawmill and is either sold as a fuel or used as fuel in the pulp mill's bark boiler. The combined steam production of the recovery boiler and the bark boiler is more than sufficient to cover the heat demand of the current processes while co-generating electricity in back-pressure turbines. Therefore, excess steam is also used for electricity generation in a new condensing turbine.

Thermal stream data for the current pulp mill was based on Pedersén and Larsson (2017) and was used with the following modifications. All streams relating to feed water heating were removed. For direct injected process steam, streams corresponding to water heating to the saturation temperature were added. Streams representing convective and radiative heat release in the recovery boiler were added and any streams relating to the operation of the bark boiler were removed, since this was considered a utility boiler.

The firing of black liquor in the recovery boiler is part of the process and the generated high temperature heat was considered unavoidable. Due to discrepancies between steam consumption and steam production in the data reported in Pedersén and Larsson (2017), the recovery boiler load was estimated by matching steam production against process steam consumption and power production data given in Pedersén and Larsson (2017). As shown below, in the case of ideal heat integration, the bark boiler is not necessary and in order to minimise fuel consumption it was assumed that all bark is exported after drying.

PROCESS DESCRIPTION: CARBON CAPTURE (BIO-CCS)

The considered CCS technology is post-combustion capture using mono-ethanol amine (MEA) as a solvent. The only significant heat demand of the capture process is reboiler steam for regeneration of the MEA-solvent, which was assumed to require 3.2 MJ/kg CO_2 at a temperature of 120 °C (IPCC 2005; see e.g. Normann *et al.* 2017). The cooling demands of the capture process include, e.g., the desorber overhead condenser and compressor intercoolers, and were estimated based on the work of Pfaff *et al.* who analysed the integration of a post-combustion capture process with a coal-fired power plant (Pfaff *et al.* 2010). No further use of the CO_2 on-site was considered in this case-study.

The main sources of CO_2 emissions at the pulp mill are the flue gases of the recovery boiler, the bark boiler, and the lime

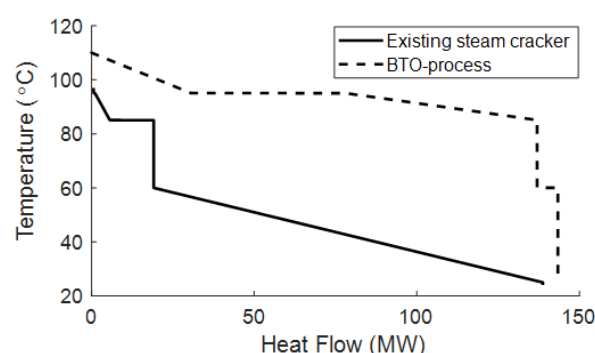


Figure 4. Comparison of excess heat signatures for the existing steam cracker and the BTO-process.

kiln. For our analysis, only the CO_2 from the recovery boiler and lime kiln were considered for CO_2 capture. The recovery boiler flue gases are relevant to consider due to the large flow, and the flue gases of the lime kiln are interesting due to their high concentration of CO_2 . The flue gases of the bark boiler were not considered due to intermittency and a relatively smaller flow. The emissions from the recovery boiler and lime kiln will not change due to efficiency improvement in the heat recovery system, since their operation is determined by the pulp production rate.

The flows and CO_2 concentrations of flue gases were estimated to be 860,000 Nm^3/h at 13 vol-% for the recovery boiler and 67,000 m^3/h at 20 vol-% for the lime kiln. Consequently, this would correspond to CO_2 emissions of around 62 kg/s from the recovery boiler, and 5 kg/s from the lime kiln, or 67 kg/s in total. In the analysis below, it was assumed that 90 % of the CO_2 is captured leading to a total heat demand for capture of 193 MW.

EXCESS HEAT EVALUATION

The split GCC of the current pulping process and the integrated steam cycle is presented in Figure 5 (left). To ensure that the current 30 MW of district heat (at 40.9/95.8 °C) can be exported from the mill in the theoretical case – according to the prioritisation order listed in the Methodology section – the LP steam extraction at 145 °C is slightly larger than required to meet process steam demand. The resulting theoretical XHT-

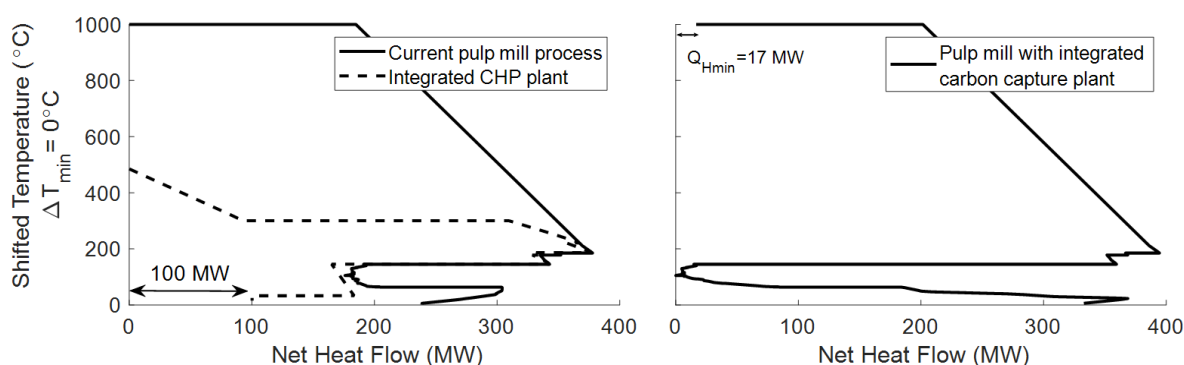


Figure 5. Left: Split GCC of the current pulp mill process (continuous line) and the integrated steam cycle (dashed line). The recovery boiler is included in the process GCC and the power generation target (100 MW) is indicated in the figure. Right: GCC of the pulp mill process including the recovery boiler and the steam demand for carbon capture. No steam cycle for power co-generation is integrated.

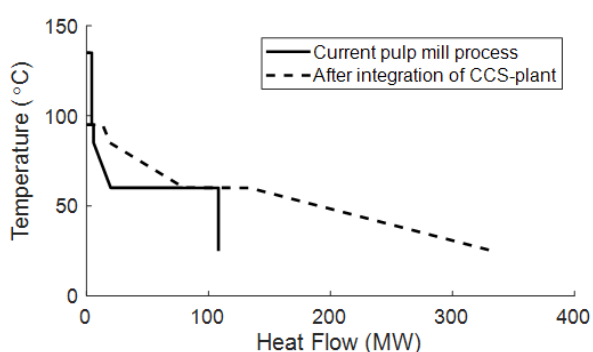


Figure 6. Theoretical XHT signature for the current pulp mill process and for the pulp mill after integration of a carbon capture plant. In both cases bark use is minimised.

signature of the pulp mill (including the pulping process and the steam cycle with a condensing stage) is presented in Figure 6 (solid line) indicating roughly 100 MW of excess heat, mostly available isothermally at 60 °C.

The very large increase in steam consumption due to the integration of the carbon capture plant (489 MW compared to 296 MW without carbon capture) obviously has major consequences for the mill steam balances, fuel consumption and power generation targets. With the prioritisation order outlined in the Methodology section, minimisation of fuel use was prioritised over co-generation. As a result, the theoretical integration of the carbon capture plant requires a complete shut-down of all steam turbines. This way the steam from the recovery boiler can almost cover the additional steam demand for carbon capture. Only a small remaining heat demand needs to be covered by the bark boiler (17 MW), see Figure 5 (right). The resulting Theoretical XHT signature is illustrated in Figure 6 (dashed line).

As shown above, the implementation of large scale carbon capture at the investigated pulp mill severely affects the mill energy balances. Significant amounts of low temperature excess heat must be cooled from the carbon capture process, and compared to the Theoretical XHT signature before the integration

of the capture plant, the amount of excess heat at or above 60 °C is increased by approximately 30 MW while almost 200 MW is added in the very low temperature range (25–60 °C). Since the heat demand of the carbon capture process is at a higher temperature than the excess heat of the current process, its heat demand is covered by using steam that previously enabled about 100 MW of electric power generation in the steam turbines.

Since eliminating cogeneration of electric power may not be an acceptable option for the mill, we also analysed the effect of changing the prioritisation order between fuel savings and co-generation, thus steering the solution towards combined heat and power production. The resulting solution involves increasing the steam production from utility boilers (e.g. bark boilers) to 127 MW. This allows for maintaining and even increasing the co-generation of electricity to 110 MW. However, excess heat levels are unaffected compared to the carbon capture case in which fuel use is minimised.

Conclusions

In this work we demonstrated a systematic methodology for studying effects of new technologies for emission reductions on theoretical excess heat availability targets for specific sites. The results of the case studies show that these technologies will change availability and quality of excess heat, as well as power generation targets, thus illustrating the importance of accounting for future process development when estimating excess heat potentials. The exact numerical results are, however, associated with many uncertainties and theoretical assumptions. While the results show how the theoretical targets change due to the integration of the new technologies, the actual future potentials will also be affected by, for example, feasibility constraints for process heat recovery and practical limitations for combined heat and power production. Nevertheless, by observing how the theoretical excess heat availability changes, we obtain an indication of the order of magnitude and direction of changes that can be expected in actual future excess heat levels.

To summarise, new technologies for reduction of industrial emissions will affect the availability of industrial excess heat if implemented on a large scale. In the case studies of the present work, the results indicate that the quantity and quality of excess heat will increase after implementation of such technologies,

and that power imports will increase significantly. For case study 1 (conversion of a steam cracker to a bio-olefins plant) increased power consumption leads to a power deficit despite an increased power generation target, while for case study 2 (integration of full scale carbon capture at a pulp mill) the power generation target reduces to zero, turning a power surplus into a deficit.

The included case studies are too few to allow for general conclusions and the results for other sites will be different depending on the technology used, the type of industry, and specific site conditions. However, if complemented by more case studies, the systematic approach applied in this study could form the basis for further development of a bottom-up methodology for generalising and aggregating results of individual case studies into regional or sector-wide potentials for future availability of industrial excess heat.

Future work could also investigate how the results are affected by changes in the prioritisation order between fuel, power and excess heat. Other potential developments include a more explicit incorporation of heat pumping opportunities for increased on-site heat recovery in the targeting method.

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