Invited Review Paper:
Process integration methodologies developed and applied in Swedish kraft pulp mills - A review, by M. Mahmoudkhani and T. Berntsson

Award-Winning Papers:
Howard Rapson (bleaching)
Douglas Atack (mechanical pulping)
& Douglas Jones (environmental)
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A Word from the Editor-in-Chief and the Deputy Editor

We have a special invited review paper in this issue of J-FOR, certain to interest our readers. Maryam Mahmoudkhani and Thore Berntsson of Chalmers University of Technology were invited to give a historical perspective on process integration and energy analysis techniques in energy-intensive industries such as the pulp and paper industry. More recently, these techniques are being adapted to address different biorefinery concepts that are emerging for retrofit to existing pulp mills – a critical issue for many forestry companies, especially in Canada where energy efficiency must be considered hand-in-hand with the implementation of biorefinery concepts. Maryam and Thore present an interesting case study in this context.

It is interesting to look back in time at energy analysis, to appreciate the knowledge basis for this paper, and the complexities of industrial processes. Process integration was established as a scientific area in the 1970s, when pinch analysis as well as mathematical programming and energy analysis for complex industrial systems emerged. Originally these methods focused on energy efficiency but have since been considerably broadened into non-energy aspects, total site analysis, GHG mitigation opportunities, methods for merging of the original concepts, etc. Process integration methods specifically for application in the pulp and paper industry were developed in the late eighties and the nineties, mainly in Scandinavia and North America. These methods specifically address the complex interaction between water usage and energy consumption as well as between process water and cooling water, feedstock for use both as product and energy, and (relatively speaking) spatial layout of energy streams in mills.

We hope you will enjoy this issue which also features some excellent papers in J-FOR’s traditional area contributions.

A Word from the PAPTAC Director

We are very pleased to be bringing you this issue of J-FOR, featuring three PAPTAC award-winning papers that were presented at PaperWeek Canada - the winners of the Howard Rapson Memorial Award for Best Chemical Pulp Bleaching paper, the Douglas Atack Award for best mechanical pulping paper, and the Douglas Jones Environmental Award for the best paper contributing to environmental improvements.

Each of the award-winning papers presented solid technological advancements and sound science in their respective fields. Furthermore, mill participation & input was significant with 2 of the 3 papers entirely written by mill personnel. I wish to take this opportunity to congratulate the winners and encourage all authors to present their best material at PAPTAC conferences and be recognized by their peers.

Erratum: Please note that there was an erratum in the previous issue (Vol. 1, No. 3) on p. 57. Equation 1 was duplicated and should not have appeared twice. We apologize for any inconvenience this may have caused.
PAPTAC Supports FPAC’s Vision 2020 - Canada’s Natural Advantage

FPAC (the Forest Production Association of Canada) recently released its industry-led Vision 2020 under the brand of “Canada’s Natural Advantage”, charting a bold and innovative future for Canada’s Forest Products Industry and outlining where it sees itself by the year 2020.

PAPTAC supports this vision that focuses on the Canadian forest products industry to power Canada’s new economy by being green, innovative and open to the world, through three major components and objectives: People - by renewing the workforce with at least 60,000 new recruits; Environment - by delivering a further 35% improvement in the sector’s environmental footprint; and Products - by generating an additional $20 billion in economic activity from new innovations and growing markets.

We are very pleased to see several key players of the industry supporting this vision which targets essential elements that will allow the industry to grow and prosper. As such, PAPTAC is committed to helping carry out this vision through our programs and activities and to collaborate with FPAC to serve our members and the industry in the path of Vision 2020.

As the technical and business association serving the industry, PAPTAC is putting in place new programs and resources addressing the need to assist and renew the industry’s future workforce. Such programs respond to the essential need for an effective transfer of acquired knowledge and know-how to the future generations of workers. The National Job fair being organized as part of PaperWeek Canada 2013, will be a great opportunity to bring together students, young workers and HR Managers on a National platform and dynamic environment.

As a leader in the dissemination of information on industry research and the improvement of operations, forest advancement is also at the forefront of our objectives. With vehicles such as J-FOR, our industry conferences as well as our Technical and Business Communities, PAPTAC provides programs and resources to support the transformation and innovation of Canada’s pulp/paper and forest products industry as it diversifies and tackles new markets.

We look forward to collaborating with FPAC and the industry to achieve these objectives and continue to build the significant progress that has been made. [www.fpac.ca/index.php/en/vision2020/]

UBC to Lead PWC Future Workforce Program

The University of British Columbia, in collaboration with PAPTAC, will lead the Future Workforce Program at the 2013 edition of PaperWeek Canada. (www.paperweekcanada.ca). In partnership with the PAPIER organization (the Canadian Pulp and Paper Network for Innovation in Education and Research), representing the main University Pulp and Paper Centres in Canada with UBC, UoF, McMaster, Lakehead U, École Polytechnique, UQTR, McGill and UNB, PaperWeek will feature an extensive program including Student Conferences and a Job Fair, focused on the Future Workforce of the industry - an increasingly important issue for the industry of tomorrow.

Update on the PAPTAC 2012 Executive Council

PAPTAC is pleased to announce the nomination of Eric Ashby, Mill Manager at Domtar Windsor, as incoming member of the Executive Council. Mr. Ashby will begin his three-year term and serve as councillor for the 2012-2015 period. We also take this opportunity to recognize Martin Fairbank who has retired from a four-year term on Council. Mr. Fairbank demonstrated great leadership and involvement during his tenure and we wish to thank him for his dedication to the Association.

New Leadership of the Process Control Community

We are pleased to announce that Holton Quinn, Process Control Engineer at Alberta Newsprint Company will be the new chairman of the PAPTAC Process Control Community. The leadership of the community met during the Control Systems Conference in New Orleans, LA to discuss future plans and projects for the coming year and is also considering holding the next edition of Control Systems (the world’s preeminent meeting for pulp and paper process measurement, control and systems engineering) in 2014 in Canada. Holt has recently been awarded the Douglas Attack Award for Best Mechanical Pulping paper, for “Mill Experience in Newsprint Quality Prediction at the TMP Refiners” presented at PaperWeek Canada 2011.

Ralph Lunn Senior Process Engineer Zellstoff Celgar Limited Partnership

Ralph is a Mechanical Engineering graduate who started his career at Crestbrook Forest industries in Skookumchuck B.C. (now Tembec) as a Project Engineer. He did this for 3 years before going back to school to complete his Masters degree in Pulp and Paper from McGill in 1992. Upon graduation, he started as a Process Engineer for Celgar Pulp in Castlegar B.C. While his work at Celgar has seen him in several different departments, including technical, engineering, operations, and management, the majority of his time has been in the technical or process group. He has always enjoyed the problem-solving aspect of being a Process Engineer, and the wide variety of learning that continues in every area. An important part of the job for him is good documentation of work done; this is evident in the technical papers he has written and presented over the years on topics such as bleach plant process optimization and washing, off-grade analysis and identification, energy improvements, and the chlorine dioxide plant.

Ralph feels that some of best opportunities to learn from others has come through various PAPTAC communities – the Bleaching Community and the Alkaline Pulping Community in particular. Connections made here have continued for many years, and have been valuable from both a technical and personal perspective. More recently Ralph became involved in the PACWEST Committee, and served as chair for the conference held at Sun Peaks in 2010. As PAPTAC mill champion, he tries to ensure that people know about and attend the various committees and conferences, as he feels they offer great value.
ABSTRACT

A wide range of investigations into process integration in energy-intensive industries such as the pulp and paper industry have been conducted by various universities and R&D centers both in Scandinavia and in North America. Various methods and tools to improve the energy efficiency of pulp and paper mills have been developed. Recently, implementation of biorefinery concepts in existing pulp mills has been the major focus for integrating these new processes into pulp mills. In this paper, an overview of methods, tools, and methodologies used by key players in the field of pulp mill-based biorefinery system studies is presented; the focus is on methods developed and applied in Swedish Kraft pulp and paper mills. The strengths and weaknesses of existing methodologies are discussed, and an example of application of newly developed methodologies to the implementation of new concepts such as biorefineries is presented.

PROCESS INTEGRATION METHODOLOGIES DEVELOPED AND APPLIED IN SWEDISH KRAFT PULP MILLS—A REVIEW

MARYAM MAHMOUDKHANI, THORE BERNTSSON*

INTRODUCTION

Improved energy management, process integration, and water reduction are among the measures that have been proposed to improve energy efficiency in the pulp and paper industry. Process integration tools and methods have been used to analyze the use of energy and other resources in this industry and to identify ways to increase productivity, decrease costs, and address environmental issues. These tools have been used to perform process-integration case studies of pulp mills from 1990 onward in various locations, including the United States, Canada, Finland, and Sweden. Projects in North America have focused primarily on possibilities for improving resource use in the short and medium term, while studies in Sweden have been more strategic in nature and have had a more long-term implementation perspective (strategic decision-making under future energy-market uncertainties) [16].

Mill specifications depend on technical, geographical, and legal parameters and differ in the degree of modernization of equipment, water and energy systems, fuel consumption, carbon dioxide emissions, and applicable taxation and policy instruments. In this chapter, the interaction between various Kraft pulp mill specifications and the choice of an appropriate process integration method is described. The Kraft process was chosen for study because the heat and water systems in Kraft pulping are more complex than those in other mills such as thermochemical pulp (TMP) mills.

Available data on energy use in the pulp and paper sector indicate an approximately 30% higher energy consumption per product unit in North America, which is due mainly to the aggressive modernization and expansion programs undertaken in Scandinavian mills and the energy optimization strategies undertaken in response to high fuel prices. The literature contains numerous studies describing various methods and tools for process integration in the pulp and paper industry. However, a summary of the methods developed and their applications in the Kraft pulp and paper industry has not yet been published. This paper aims to provide an overview of the available methods and tools which have been used for process integration in the Swedish Kraft pulp and paper industry.

Various methods are described and discussed, and a few examples of applying these methods to mill energy analysis are presented.

PROCESS INTEGRATION STUDIES

The term process integration refers to a systems-oriented and integrated approach,

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meaning that rather than optimizing process units separately, the interaction between different parts of the process is considered, and the system as a whole is optimized. Process integration studies can be based on mathematical methods, thermodynamic methods, economic methods, or a combination of these. Examples of these methods include mathematical programming, pinch analysis, and hierarchical analysis. The most widespread methods for process integration in the pulp and paper industries are pinch analysis, mathematical programming, and to a certain extent exergy analysis, among which pinch analysis methods are well proven and established and the most commonly used in commercial projects (IEA/Annex XIII, 2004).

In general, exergy analysis is used when there is a large pressure difference in a process unit or for electricity production units, particularly when electricity and heat production interact. Exergy analysis has, however, been used only sporadically in the pulp and paper industry because electricity and heat are always treated separately in energy balances. Exergy analysis has, however, been used in combination with pinch analysis to provide a new perspective on the identification of process enhancement opportunities involving energy upgrading and conversion in the pulp and paper industry. Mathematical programming, on the other hand, is a tool by which mathematical relations can be identified in the system under analysis. The MIND (Method for analysis of INDustrial energy systems) tool developed in the Department of Management and Engineering in Linköping University (in Sweden) is an example of a tool based on mathematical programming methods which has been applied to the analysis and optimization of industrial energy systems as well as to industries in cooperation with other players such as district heating systems. To the authors’ knowledge, there exist very few studies of the application of mathematical programming methods to energy analysis of Kraft pulp mills [19,18,10]. No details of these methods are therefore discussed in this paper. The focus of this paper is therefore on the methods developed based on pinch analysis.

An existing concept based on pinch technology and used for performing retrofits or for integrating new energy technologies is the grand composite curve (GCC). Pinch analysis tools are used to construct GCCs, which can then be used to analyze a mill’s energy systems. Early examples such as TARGET (written at UMIST) and PROTAB (developed by ICI) have been developed into highly sophisticated programs such as SUPERTARGET (Linnhoff Marxl) and Advent/Aspen Pinch (Aspen Tech). Later, Hyprotech developed HXNET, which became Aspen’s main specialized pinch software module. These programs use complex targeting procedures and are relatively costly to purchase or license [20]. Tools such as Pro-Pi (add-in module in MS Excel developed by CIT, Chalmers Industrieteknik) have therefore been developed to perform the essential calculations of energy and cost targets and to generate GCCs without including the network design. Pro-Pi has been used for research purposes and in applied industrial studies in Sweden and Finland. Canmet ENERGY (http://canmetenergy-canmetenergie.nrcan-rcan.gc.ca) recently developed and tested a process integration software tool called “INTEGRATION” in Quebec. The tool aims to make process integration studies easier, faster, and cheaper and focuses on thermal energy usage (i.e., energy performance, quantity and quality of waste heat, and opportunities for heat recovery). The “INTEGRATION” tool is not yet completely documented, but as a continuation of the research work reported here, it could be interesting to study the possibilities of combining this tool with the recently developed methods presented in this paper.

Pinch analysis: a thermodynamics-based method

Pinch analysis is a method with a structured approach which is used to design energy-efficient systems, i.e., to maximize internal heat recovery within a process and to minimize process demands for the hot and cold duties supplied by utilities [25,40,20,23]. By knowing the heat content of all streams and their start and target temperatures, a heat-exchanger network (HEN) can be designed in which the external heat demand is minimized. It has, however, been reported that the traditional pinch method is ill-suited to tackle certain problems. For example, pinch curves are not well suited for retrofit situations because no information on the existing HEN design can be extracted from these curves [32]. In addition to the lack of information on the existing HEN in traditional pinch analysis, neither the direct relation between energy and water reduction nor the relation between the use and production of utilities is considered. Because water handling in the pulp and paper industry is complex and closely related both to energy use and to the environmental impacts of the mill, other approaches have been developed to tackle these issues. In the literature, various kinds of systematic methodologies combined with thermal pinch analysis have been presented. Examples of such methodologies are the combined method used in the “NRCAn” optimization software package used by VTT in Finland and the SDM™ (“Successive Design Methodology,” developed by American Process, Inc., provides a systematic approach for simultaneous application of water and energy conservation in practical mill situations) software package (IEA/Annex XIII 2004).

ENERGY-EFFICIENCY MEASURES BASED ON NEWLY DEVELOPED METHODS

There are numerous pathways for decreasing energy usage in pulp and paper processes. Overviews of various measures to save both thermal and electrical energy can be found in [8,9,27]. In this paper, the focus is on measures used to achieve thermal energy savings, such as steam savings and excess heat utilization. The main method for identifying efficiency measures described in this paper is process integration (so-called heat integration), which can be carried out with or
without major changes in process equipment. The measures most frequently mentioned in the literature are internal heat recovery, condensate recovery, redesign of the hot- and warm-water systems (HWWS), and water reuse, as well as energy conversion and upgrading of process equipment, e.g., process-integrated evaporation units [32,45,46,16,47,5,29,30,14].

It is important to note that when performing energy analysis in a mill, the best approach is not always to begin by eliminating or reducing the pinch violations and then to use the remaining excess heat to reduce energy consumption. In many cases, efforts to eliminate or reduce pinch violations may encounter practical constraints, may be economically unattractive, or both. The most cost-effective solution with respect to how many, and which, inefficiencies (i.e., pinch violations) to resolve is actually an optimization problem. The target function is the net profit of the mill. For strict mathematical optimization, conditions such as the mill layout and piping distances are needed. Moreover, economic conditions such as energy prices and the annuity factor are also needed to find the best solution [4]. To evaluate the future energy export potential of a mill, both energy prices and the CO2 emissions impact of energy exports and imports would be needed. Therefore, the need to develop new methods to provide the required information is obvious. This topic is discussed further in detail later in this paper.

Pinch-analysis-based advanced composite curves

The grand composite curve (GCC) is not well suited for retrofit situations based on common pinch analysis in the following situations [31]:

- In GCCs in which the temperatures are not estimated correctly in existing systems having a heat demand that is large in relation to what could be supplied internally within the process. The GCC may not be able to capture the full potential for heat recovery.

**TABLE 1** Description of advanced pinch curves [31,32].

<table>
<thead>
<tr>
<th>Hot Utility Curve (HUC)</th>
<th>Actual Heat Load Curve (AHLC)</th>
<th>Theoretical Heat Load Curve (THLC)</th>
<th>Extreme Heat Load Curve (EHLC)</th>
</tr>
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<tbody>
<tr>
<td>The HUC represents a composite of the hot utility as a temperature versus heat load diagram (e.g., for water, steam, or oil distribution systems).</td>
<td>The AHLC represents a composite of all the cold streams or cold stream segments constituting the cold side of the heaters in the heat distribution system(s) in a temperature versus heat load diagram.</td>
<td>The THLC represents the theoretically lowest temperature versus heat load at which heat can be introduced to the system when all thermodynamically correct opportunities for heat exchanger enlargement and introduction of new heat exchangers have been implemented at the existing external heat demand.</td>
<td>The EHLC represents the heat load vs. temperature level relationship when the heat exchange is carried out with a minimum heat-exchanger area. This is the same as performing the heat exchange vertically in the composite curves. It also means that the externally supplied heat is transferred to the cold streams at as high a temperature as theoretically possible.</td>
</tr>
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</table>

In principle, the same four types of curves should also be constructed below the pinch and are called the cold utility curve (CUC), the actual cooling load curve (ACLC), the theoretical cooling load curve (TCLC), and the extreme cooling load curve (EHLC). With the aid of these curves, three levels of complexity of changes in heating, cooling, and process systems can be identified and evaluated. These are:

**Improvements involving only HUC/CUC:**

Only the utility system is involved. In principle, these have no influence on the flexibility, controllability, or operability of the plant. Little or no investment cost in the heating, cooling, or process systems is required.

**Improvements involving AHLC/ACLC:**

Changes in the heating and cooling distribution systems are needed (e.g., adding one or more temperature level(s) to the heating or cooling system). No changes in the process heat-exchanger network (HEN). Relatively minor influence on flexibility, controllability and operability. Moderate investment cost in the heating, cooling, and process systems.

**Improvements involving THLC/TCLC:**

Changes in the HEN are needed (new units or enlargements of existing units). Risk of major influence on flexibility, controllability, and operability. Risk of considerable investment cost in the heating, cooling, and process systems.
be achieved by enhanced heat recovery.

- In GCCs in which no information about the existing heat-exchanger network (HEN) is provided, which means that no indication of the changes needed in the HEN to approach theoretical levels can be given.

One way to address these issues is to apply the proposed advanced composite curves developed in the Division of Heat and Power Technology, Chalmers University of Technology. Table 1 provides a description of the advanced composite curves and presents how the different levels of complexity of changes in heating, cooling, and process systems can be identified. The advanced composite curves are based on pinch analysis and provide an accurate representation of existing loads as well as possible improvements that could be obtained by retrofitting. As emphasized above, the current status and existing design of the energy systems in a plant can influence considerably the available opportunities for improvement [31]. Figure 1 illustrates and describes the different composite curves above and below the pinch point in a temperature-heat load diagram. Using these curves, important information on the various potentials of a number of different measures, such as improved heat exchange, introduction of a heat pump or a CHP plant, and the consequences of various process changes, can be extracted. These curves can therefore be used to assess possible heating-system improvements. A more comprehensive description of the method can be found elsewhere [34,35].

Generally speaking, with any energy improvement measure, there is a trade-off between the possible magnitude of the energy improvement (normally influenced favorably by greater complexity of heating, cooling, and process system changes) on the one hand and investment cost, flexibility, controllability, and operability on the other. Therefore, it is of great importance for improvement as a function of their level of complexity [32].

**“Matrix method” for economical retrofit of a heat-exchanger network**

When retrofitting an HEN using the traditional pinch method, many important parameters such as heat-exchanger area, number and type of heat exchangers, physical distance between the streams, piping, annual pressure-drop costs, and others are not considered. An approach is therefore desirable which could be used to identify many possible heat-exchanger alternatives for HENs. In this approach, it is important that the area cost not be the only cost considered, because in practice it accounts for only a fraction of the total retrofit cost. A proposed method which systematically and rigorously solves the problem of determining retrofit solutions is the matrix method [11]. In this method, the complex heat-exchanger network design has been simplified using certain heuristics such as temperature differences in the heat exchangers, tick-off matches, and stream splitting. The rearrangement of heat exchangers and thus the identification of profitable solutions are carried out in a simpler way in this method than in more advanced mathematics-based methods (e.g., MILP and MINLP as presented in [48] and [12]), and the final solution is found by comparing the alternatives identified.

In the matrix method, a matrix is developed in which hot streams run vertically and cold streams horizontally. For each thermodynamically possible match, the cost of implementing the match is calculated at a given level of energy recovery (i.e., $\Delta T_{\text{min}}$), using all costs discussed above. In each match, at least one stream is ticked off. For piping, the geographical distances between all streams are included as input data. The matrix method is an interactive method in which the user decides which heat-exchanger combinations will be included in the solution. The combination of matches that yields the lowest cost is the solution to the retrofit problem at each global $\Delta T_{\text{min}}$ [32]. The matrix is then fully recalculated with the ticked-off stream(s) in the chosen match omitted. This goes on until all matches have been chosen and the given energy recovery has been achieved. This procedure must be repeated once or twice until the sequence with the lowest total cost has been found. This method has been used successfully to find an economically feasible heat-exchanger network retrofit for the release of excess heat discovered using the curves [7].

The whole procedure described above can be carried out also with other $\Delta T_{\text{min}}$ values. This will lead to a graph of total investment cost versus energy recovery level, from which the optimal level of energy recovery for a given energy cost can be determined.

**Combined water and energy reduction analysis**

Classical water-pinch methodologies have proved to be difficult to apply in the pulp
and paper industry because there are too many contaminants in the water streams and because water is both a utility and a process stream in this industry [16,24,38]. Modern energy-efficient mills often have low water consumption, and traditional heat-recovery projects suggested mostly for earlier mills are often not applicable to modern mills [44]. Approaches for combining heat and water reduction techniques have therefore been developed in which modified water-pinch approaches have been applied in a few studies considering property-based design problems such as non-process elements, pH, density, Basic Oxygen Demand, Chemical Oxygen Demand, and toxicity in addition to water and energy issues [6,24,7,13,39,28,30,21].

In general, the water that is used as a hot or a cold utility in a mill is stored in tanks which are normally at three to five different temperature levels. A given water stream normally does not have the same temperature as the overall tank; this means that a mixing of waters at different temperatures occurs in tanks. For modern pulp mills in which the water consumption is low and the pinch temperature is already at 100°C or above, mixing of water streams at different temperatures in tanks is not considered to be a pinch violation (temperature levels are below the pinch temperature), and various solutions are required principally for improving energy efficiency. The “tank curves” method was therefore developed to maximize the usable excess heat potential at high temperature in the HWWS which can be used for other processes or saved as primary heat [33]. In this method, the system boundary encloses the HWWS, and making changes in the HWWS would then leave the remaining process without any changes at all. Therefore, no complex simultaneous optimization and analysis of water consumption, water contamination, number of tanks, temperature levels in tanks, and tank size is performed in this method. Instead, the number of tanks and their temperature levels are optimized for a given mill water demand. The method can be used in mills with either high or low water consumption.

**EXAMPLE OF APPLICATION OF ADVANCED PINCH CURVE IN BACKGROUND / FOREGROUND ANALYSIS**

When steam is saved and excess heat is released using one or more of the energy-efficiency measures mentioned earlier, this heat can be used in various ways in the mill; for example, it can be used internally to increase the level of efficiency in the mill, e.g., upgrading the evaporation plant, or externally by the installation of new equipment like heat pumps, or in the implementation of new biorefinery concepts, e.g., precipitation of lignin from black liquor. An example of using advanced pinch curves in implementing a biorefinery concept at a Kraft pulp mill is presented below.

For analyzing the energy consequences when a sub-process, e.g., a distillation plant, is changed or when a new sub-process is introduced into an existing process, the concept of background/foreground analysis is useful. The concept should be seen mostly as a guide to finding the integration possibilities between processes. In this analysis, the existing process and its GCC are called the “background process” and the “background curve” respectively. Correspondingly, the sub-process and its GCC are called the “foreground process” and the “foreground curve”. If the temperature levels match, the two processes can be integrated so that the cooling demand of one process is used as heat in the other. The possibilities can then be analyzed and approximately quantified with the aid of the background and foreground curves.

One example of these advanced curves is taken from a project in which converting a Kraft pulp mill into an ethanol plant and the associated heat integration challenges have been investigated, as illustrated in Fig. 2 [14].

The challenge in this converting process is how to integrate the new sub-processes of distillation and evaporation in the ethanol plant into the background process (defined as the pre-treatment, the ethanol line (except for distillation), and the units in the chemical recovery cycle, while the foreground process is the distillation and the evaporation units). The background process studied is in part a retrofit design (the reused pulp-mill unit operations) and in part a green-field design (the ethanol...
line). The base case for comparison in this study is a background process design in which no energy-efficiency improvements have been made in the repurposed pulp mill. Advanced pinch curves, shown in Fig. 3, have been used in process integration studies for this process [14].

In Fig. 3, the AHLC (Actual Heat Load Curve) and the ACLC (Actual Cooling Load Curve) illustrate the temperature levels at which hot and cold utilities are needed in the base-case process. As can be seen from the ACLC, there is not much excess heat available for integration of the thermal-separation units in the original design (approximately 3.4 MW above 90°C). If a GCC were constructed for the process at the ΔT_{min} corresponding to the same utility demand, it would appear that the excess heat available would be 13.7 MW (above 90°C). This excess heat can, however, be made available in the process only if a retrofit of the heat-exchanger network is done. The demand for low-temperature hot utility (below 110°C) is 13.5 MW, and because the THLC and the AHLC are almost completely superimposed, the supply temperature cannot be decreased further without violating the pinch conditions.

There are two ways of studying the possibilities for steam savings in this process; either the background process could be improved by increasing internal heat recovery (i.e., resolving the pinch violations), or the heat-exchanger network could be redesigned to increase the possibilities for external heat integration by means of distillation and evaporation. From Fig. 3, it can be seen that resolving pinch violations will decrease the hot utility to 15 MW (where the THLC and the TCLC cross). It can also be seen that approximately 17 MW of excess heat at a high temperature (above 100°C) can be freed without changing the hot-utility demand (TCLC). The theoretical cooling load curve (TCLC) shows the highest temperatures at which coolers can be placed in the process for a specific ΔT_{min} (set at 7°C in this study) while maintaining the base-case hot-utility demand.

The figure also indicates that there is a pinch violation of approximately 2 MW which can be resolved with the excess heat existing in the process today (to the left of where the ACLC and AHLC cross). This pinch violation actually represents the use of low-pressure steam (4 bars) for building heating. The total of the pinch violations that can be resolved is therefore 17 MW.

**ASSESSING PROFITABILITY AND CARBON BALANCES OF ENERGY INVESTMENTS AND EVALUATING COMMERCIALIZATION OF BIOREFINERIES**

Because energy use in the pulp and paper industry is high, energy savings through process integration in this sector can have a large impact on national CO₂ emission levels; in fact, the potential reduction in GHG emissions is one of the most promising drivers for the use of process integration tools in this industry [26,22,44,43]. The profitability and the net CO₂ emissions reduction potential of energy investments in this sector must therefore be assessed by quantifying their implications within a future energy-market context. To carry out decision-making subject to uncertainty about future energy-market conditions, various strategies can be considered.

Over more than a decade, research on strategic methods for evaluating candidate investments, using different scenarios that include future fuel prices, energy carrier prices, and CO₂ emissions and policy instruments, has been conducted in the Division of Heat and Power Technology at Chalmers University of Technology [1-3]. A calculation tool, ENPAC (Energy Price and Carbon Balance Scenarios, developed by Erik Axelsson and Simon Harvey at Chalmers University of Technology and Profu - www.profu.se), has been developed for creating consistent scenarios adapted for this energy-intensive industry as well as for obtaining reliable benchmarks [4]. The ENPAC tool calculates energy prices for a large-volume customer based on forecasted world-market fossil-fuel prices and relevant policy instruments (e.g., costs associated with emitting CO₂, various subsidies favoring renewable energy sources in the electricity market or the transportation-fuel market) and on key characteristics of energy-conversion technologies in the district-heating and electric-power sectors. An overview of the procedure and purpose of the ENPAC tool for evaluation of energy-efficiency investments in energy-intensive industries is shown in Fig. 4. For further details, the reader is referred to [3].

A systematic approach which trades off CO₂ emission reductions against investments in process-integration measures in industry while optimizing these

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Fig. 3 - Advanced pinch curves for the background process [14].
investments under economic uncertainty has also been developed in the same division [41,42]. The methodology proposed in [42] is based on the assumption that investment decisions must be made before the values of uncertain parameters can be known. Many of the uncertainties affecting investment decisions are related to future CO₂ emission targets and policies. The optimization should therefore consider not only economic criteria, but also greenhouse-gas reductions. As the decision problem became a multi-objective programming problem, the method proposed by Svensson and Berntsson attempted to provide a good framework for decision-making about energy-efficiency measures when considerations of greenhouse-gas reductions are influencing these decisions.

The uncertainties affecting decision-making for implementing biorefinery concepts is not limited to decisions about energy-efficiency measures. Commercialization of biorefinery processes may still take some years due to a wide variety of barriers and risks, e.g., future access to raw material at a competitive price [17]. To mitigate these risks in design decision-making, technical- and market-based uncertainties should be considered in techno-economic assessments, using, for example, Monte Carlo analysis to account for these risks in preliminary screening of integrated forest biorefinery options [15]. It is also of crucial importance that the profitability of different possible production routes and product portfolios be evaluated while stakeholder value is maximized through global optimization of the supply chain. Tools based on mathematical optimization have been developed to achieve these ends [37]. Other methodologies which have been developed and presented in the literature provide a systematic way to synthesize and screen quickly a number of biorefinery pathways as both technology- and optimization-based frameworks to reduce the number of possible pathways from the synthesized ones without missing the globally optimal solution [36].

SUMMARY

This paper was intended to provide an overview of the available thermodynamics-based methods and tools that have been used to perform process integration studies for Kraft pulp and paper mills in Sweden. This overview could provide a basis for building a platform for further development of process integration tools and methods, including consideration of the implementation of new biorefinery concepts. In this context, a number of various approaches which tackle the complex issue of process integration with regard to energy and water systems, appropriate efficiency measures, and evaluation of the chosen integration measures under an uncertain future energy market have been presented. Depending on the operating history of the mill under study (modern vs. old mill) and the degree of existing integration, different methods can be used to optimize process energy consumption. In general, it can be said that energy savings can enable energy exports from the mill or can be used to integrate new technologies or concepts into the mill. To decide which measures to choose, a method for assessing profitability and CO₂ balance has been presented.

ACKNOWLEDGEMENTS

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REFERENCES


34. Nordman, R. and Berntsson, T., “Use of Advanced Composite Curves for Assessing Cost-Effective HEN Retrofit: Theory and...
EMERGING AREA CONTRIBUTIONS

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ABSTRACT

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Many pulp mills that have adopted the ECF (elemental chlorine-free) bleaching process have experienced various degrees of inorganic scale deposition in their bleaching equipment. Typical scales found after this conversion are calcium oxalate (CaC\(_2\)O\(_4\)), barium sulphate (BaSO\(_4\)), and calcium carbonate (CaCO\(_3\)). The primary sources of these deposit constituents are the incoming wood supply, especially bark, mill water, and ionic species produced by the pulping and bleaching processes. This paper presents the case of one mill in which this scaling caused operational, production, and quality issues. A method was developed to identify and monitor incoming calcium levels to the bleach plant. This method and others were used to evaluate an anti-scale control program and its impact on operations. Mill benefits along with those of washer/pipe inspections are demonstrated.

INTRODUCTION

Adoption of the ECF bleaching process has caused some changes in the bleach plant, especially in the first bleaching stage. Essentially, the ECF process completely replaces chlorine (Cl\(_2\)) by chlorine dioxide (ClO\(_2\)). The latter chemical is a powerful oxidizer and offers extensive benefits for pulp fibre bleaching. However, one detrimental aspect of ClO\(_2\) bleaching is that it does not generate as much HCl (hydrochloric acid) as does chlorine during the bleaching reactions. Therefore, the end pH in the D0 stage (first ClO\(_2\) stage) will be slightly higher when using 100% ClO\(_2\). At the DMI Peace River Pulp Division (PRPD), the D0-stage pH (washer feed pH) was increased from nearly 2.0 to 3.0 after ECF conversion. At higher pH, metals will tend to be less soluble and prone to precipitate. These changes will affect the stability of these metals, and scale deposition may occur, which leads to plugging in washer decks, pumps, pipes, and other locations. This scenario also allows more metals to be carried forward to the extraction stage, resulting in rapid deposition of calcium carbonate. Many problems may arise in the bleaching process, including quality issues, production losses, and downtime, depending on the rate of precipitation and the type of scale.

It is important to mention that one simple approach to minimize overall scale formation in bleaching processes is reduction of the D0-stage pH to 2.0 or lower. In the case of PRPD, this approach would incur significant additional chemical costs (higher acid and caustic usage), and therefore an anti-scale control program was implemented and evaluated.

The main types of inorganic salts or scales found in the bleach plant are calcium carbonate, calcium oxalate, and barium sulphate. These are also the main type of scales present in the bleaching process at PRPD.

Calcium Carbonate

This salt is the most common scale found in the bleaching process. If conditions are favourable, it will rapidly precipitate and accumulate on many surfaces. Its solubility decreases with increased temperature (inverse solubility), and it also exhibits low solubility under alkaline conditions (it precipitates when the pH is higher than 8). At PRPD, if calcium levels are high in the extraction stage, then some calcium carbonate precipitation may occur.

Calcium Oxalate

This scale is commonly found in ECF mills. At higher pH (>3), oxalic acids tend to dissociate to form bioxalate and oxalate ions. If calcium levels entering the bleach plant are high, then oxalate ions may form calcium oxalate, and deposition may be observed. Note that the amount of oxalic acid found in bark is substantially higher than in wood chips.

In the PRPD mill, calcium oxalate is found mostly in the D0 stage, with some deposits in the D1 and D2 stages (second and third ClO\(_2\) stages). It is frequent to find this scale plugging the D0 washer deck, which can eventually result in high

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calcium carryover to the extraction stage.

Barium Sulphate

This scale is commonly found in the D0 stage. It can plug washers, decks, impellers, mixers, and other equipment. It is difficult to clean and can potentially cause extended downtime. At higher pH (≥3), the sulphuric acid will dissociate into bisulphate and sulphate ions, which may in turn be converted to barium sulphate. Use of sodium sesquisulphate (Na₂H(SO₄)₃) for pH control in the bleach plant is not recommended because it will increase the concentration of sulphate ions in the filtrates.

In the PRPD mill, a chelant wash boil-out (using ethylene-diamine-tetraacetic acid, or EDTA) is usually performed every three months to prevent washer-deck plugging and to remove barium sulphate and calcium oxalate.

Bark and Anti-scale Treatment

Bark is the external layer that surrounds wood, branches, and roots and can account for 10%–15% of total tree weight. Bark is also another important variable in the formation of scale during bleaching processes.

In certain wood species, bark contains 15–30 times more calcium than wood [1]. This could result in ~40% more calcium being fed to the fibre line if only 2% of the wood chips were replaced by bark.

It is known that during the cooking and washing processes (under alkaline conditions), most of the calcium is bound to the fibres and carried over to the bleach plant [2] unless an acidification stage is used. In other words, the washing process will have a minimum impact on removing calcium from the fibres, independently of whether washing is “good” or “bad.” Other variables that affect calcium levels entering the bleach plant are white liquor quality and water hardness.

In the PRPD mill, various programs have been implemented to control white liquor quality. Periodic acid washes are performed on the white liquor pressure filter to maintain optimal operations. Another measure taken was to use contaminated hot-water condensate in the last brownstock washer. This condensate is considered to be virtually free of calcium.

Further experiments performed at PRPD showed that the major contributor to higher calcium levels entering the bleach plant is essentially bark, even when all the contaminated hot-water condensate is replaced by hot mill water.

An anti-scale program was proposed and implemented at PRPD to minimize and control scale formation during the bleaching process, especially in the D0 and E0 (oxygen extraction) stages. The main benefits expected from this program were lower Cl₂ and NaOH (sodium hydroxide) usage; better pulp quality; and less downtime due to unexpected outages for scale cleaning.

The anti-scale chemical is a high-molecular-weight anionic polyelectrolyte that inhibits scale crystal formation. The inhibition is achieved by changing the surface charge of the salt crystals, thus preventing crystal agglomeration and growth. The anti-scale chemical can also modify the structure of the salt crystals, preventing them from agglomerating or aggregating.

EXPERIMENTAL METHOD

Calcium Levels

The main intent of this research was to propose a practical testing method to identify and analyze scale deposits in the bleach plant and thereby to evaluate the efficiency of the anti-scale control program.

The experimental procedure to determine calcium levels in the pulp involved measurement of sample consistency, acidification, mixing, filtering, and concentration measurement using an atomic adsorption spectrometer analyzer (Atomic Adsorption Analyzer Varian Model Spectra 55).

The usual consistency for measurement was approximately 1.0% solids, but actual consistency measurements were made on each sample using TAPPI standard method T 240 om-02. Afterwards, the samples were acidified to a pH lower than 1.0 using 93% sulphuric acid solution. Then the samples were mixed for two minutes to ensure proper acid dispersion onto the fibres. Next, the samples were filtered using glass fibre filters (47-mm previously weighted filters for total suspended solids), and finally the samples were diluted and prepared for analysis in an atomic adsorption spectrometer analyzer (AA analyzer).

The main reason to acidify the pulp fibres is to liberate calcium from the fibres. In acidic environments, calcium will be widely dispersed in the filtrates (the water phase). On the other hand, in alkaline environments, most of the calcium will be bound to the fibres [3]. By analyzing the same pulp filtrates using the procedure described earlier, it was determined that the acidified sample contained on the order of 20 times more calcium than the non-acidified sample. The numerical results were 618 ppm calcium for the acidified sample versus 38 ppm calcium for the non-acidified sample.

Roles of Washing, Carryover, and Bark

In some cases, pulp mills have replaced the dilution water in the last brownstock washer with contaminated condensate from the evaporators [4]. This condensate stream has a very low calcium concentration. The replacement of regular hot mill
water (tap water) by this filtrate should theoretically reduce the calcium load to the bleach plant. However, the results obtained (Fig. 1) indicate a minimal impact on the calcium load to the bleach plant. The samples analyzed were obtained from the last brownstock washer (roughly ~25% consistency) and diluted with their respective filtrates (tap water and contaminated condensate) to medium consistency (~10%).

The intent here is not to influence the decision whether to use contaminated condensate in the last brownstock washer, but to recognize the minimal impact of the contaminated condensate on the overall calcium load to the bleach plant. The calcium concentration in tap water was approximately 38 ppm, while that of the contaminated condensate was approximately 3 ppm.

The impact of another important variable, carryover, on calcium levels going into the bleach plant was also investigated. The results found at PRPD support the argument that most of the calcium will be “bound” to the fibres in alkaline environments, independently of brownstock washing efficiency. At PRPD, during “upset conditions” when brownstock washing was limited, the amount of calcium going to the bleach plant remained virtually constant.

Bark is the major contributor to any variation in calcium levels going into the bleach plant (assuming good white liquor quality). Figure 2a shows that throughout a six-month period, the calcium levels measured in the bleach plant at PRPD were strongly correlated with bark content. This indicates that most of the variation in calcium concentration, which can lead to scale formation in the bleaching stages, is attributable to bark content. The bark content measurement used here is a single manual test on the chips being fed to the digester during hardwood runs. This is a regularly performed test at PRPD and is based on visual inspection and the ratio of the total dry weight of bark observed [5].

Trial Results

The main intent of this trial was to evaluate the performance of the anti-scale control program. The results were primarily based on D0 washer and Eo discharge line inspections, chemical usage (ClO2 and NaOH) during the trial, effect of dirt (i.e., carbonate scale) on pulp quality, and calcium levels in the various bleaching stages.

The anti-scale chemical was applied at two different locations in the fibre line, using a total dosage of 0.3 kg/ton. The main dose was applied upstream of the MC pump before the brownstock high-density storage tank. The second location was the D0 filtrate line used for dilution. This line runs from the D0 washer to the bottom of the high-density storage tank. The main reason for choosing these locations was the hypothesis that the anti-scale should be added before the pH change, which results in liberation of calcium from the fibre surface. This approach should inhibit any crystal agglomeration or scale formation.

The calcium levels observed before and after the trial showed an improvement in calcium removal in the D0 washer. Figure 3 shows better metal removal efficiency in the first bleach washer. The samples shown in Fig. 3 were taken during hardwood runs with constant bark content.

Pulp quality was another important criterion used to determine whether or not application of the anti-scale chemical had been successful. Before the trial, PRPD had experienced some dirt outbreaks involving the presence of dirt “scale” (calcium carbonate) in the final pulp. Those episodes represented an issue for the mill because the pulp produced during that time might be considered “off-grade.” With the implementation of the anti-scale program in the bleach plant, the dirt levels decreased dramatically, as can be seen in Fig. 4. The amount of dirt scale in the final pulp was diminished mainly because of the anti-scale application. During the trial, the D0 washer stayed clean for a longer period of time, which enabled the washer to remove metals (especially calcium) efficiently, resulting in less generation of calcium carbonate in the Eo stage. The “better” washing also helped to diminish the formation of “pitch” in the extraction stage because less calcium was available for reaction.

Another factor considered was the overall chemical usage in the bleach plant, especially ClO2 and NaOH charges. Figure 5 shows a clear relationship between the total Kappa factor and the final pulp ISO
% brightness. Three cases were considered: “poor” D0 washer efficiency (deck plugging), “good” D0 washer efficiency (clean deck), and the anti-scale trial. Note that the overall ClO₂ chemical charge was slightly lower during the anti-scale trial than in the “good” D0 washer efficiency case.

Using the same method as shown in Fig. 5, the usages of caustic for the hardwood runs were compared. Figure 6 presents NaOH usage divided by bleach feed Kappa. The intent of using this ratio is to normalize the chemical usage by the incoming bleach feed Kappa. Note that the chemical usage during the anti-scale trial is lower than for the “good” D0 washer, which can represent savings of nearly 2 kg/ton of NaOH. These results support the theory that a lower chemical charge in the Eo stage is the result of a cleaner D0 washer (a clean deck). If the caustic demand is lower, then this might indicate less carryover coming from the D0 washer or a less “acidic” pulp going to the Eo stage. It is important to mention that the extraction stage at PRPD is controlled by a pH loop.

Ultimately, visual inspection of the D0 washer strongly confirmed that the anti-scale chemical (polyelectrolyte) is preventing the formation of calcium oxalate and barium sulphate in the first chlorination stage, thus keeping the D0 washer “clean.” Figure 8 shows a series of pictures of the underdeck of the D0 washer. The D0 washer is a compaction baffle filter, which is an enclosed pressurized washer with channels underneath the deck through which the filtrate flows. If these channels are plugged or “blocked,” then the majority of the pulp will not get “washed” (its filtrate replaced by cleaner filtrate), leading to “dirtier” pulp being carried through to the next stage. If high levels of carryover occur, then formation of calcium carbonate may also be observed in the Eo stage, along with higher chemical consumption.

During a 121-day period, several inspections were performed of the D0 washer using a boroscope to take pictures inside the underdeck channels (Boroscope Model Video Probe XL, Everest VIT). It can be seen in Fig. 7 that during the trial, no major signs of encrustation could be observed inside the underdeck channels. The bottom photo in Fig. 7 shows for comparison an underdeck channel completely plugged with calcium oxalate, barium sulphate, or both.

Inspections of the Eo-stage discharge line were also performed. These inspections would detect any rapid formation
of calcium carbonate in that particular alkaline stage. If high levels of carryover are observed leaving the D0 washer, then a large amount of calcium can be carried through and can react with the carbonates present in that stage. Figure 8 shows two pictures from the same process location, the Eo discharge line. The first picture, taken during a major shutdown, shows an enormous calcium carbonate formation. Such rapid scale formation is due mainly to poor D0 washer efficiency. In the second picture, calcium carbonate deposition is minimal, which can be attributed to the excellent D0 washer efficiency associated with the anti-scale trial.

**CONCLUSIONS**

The calcium concentration testing method developed in this research has been demonstrated to be an efficient tool for evaluating and monitoring the amount of calcium entering the bleach plant or any particular bleaching stage. The bark content of the chips has been shown to be the major contributor to variations in calcium entering the bleach plant and therefore the major contributor to formation of scales such as calcium oxalate and calcium carbonate.

The anti-scale application program was proven to be an efficient tool to minimize scale formation and aggregation in the bleach plant. This polyelectrolyte proved to be an ideal way to keep the D0 washer “clean” from calcium oxalate and barium sulphate build-ups.
The combination of these measures led to outstanding benefits, including lower calcium carryover to the Eo stage, resulting in less calcium carbonate formation (better pulp quality); lower ClO₂ and NaOH usage in the bleach plant; and less unscheduled down time for cleaning encrustations (lower production losses).

ACKNOWLEDGEMENTS

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REFERENCES


MILL EXPERIENCE IN NEWSPRINT QUALITY PREDICTION IN THE TMP REFINERS

HOLT QUINN

ABSTRACT

Detailed fibre size measurements from a new on-line pulp quality analyzer have helped to improve understanding of the relationship between pulp quality and paper quality. Good predictions for several newsprint properties, including MD tensile energy absorption (TEA), were developed from on-line pulp quality data from upstream of the latency chest. Those predictions were used successfully for over a year. They enabled operating and technical personnel to focus troubleshooting efforts in the correct area: the pulp mill or the paper machine. The predictions were also valuable for deciding when to change refiner plates on the three main refiner lines.

INTRODUCTION

Freeness, fibre length, and shives are typical TMP pulp quality measurements both in the laboratory and on-line. At Alberta Newsprint Company, these measurements have been successfully used to control the TMP mill operation and have been qualitatively linked to paper quality. Ideally, they would have been quantitatively linked to paper quality; however, previous attempts to do this had been unsuccessful.

In August 2008, Alberta Newsprint Company replaced their on-line PQM 400 pulp quality analyzer with a Kajaani-MAP analyzer which included Freeness and Fibre-Shive modules. Using image analysis, the Fibre-Shive module provides improved fibre size information as well as average fibre length and percent shives. The improved measurements, along with subsequent discussions with consultants in early 2009, prompted a new attempt to predict paper properties from pulp quality. Some articles authored by Olof Ferritsius [1–3] also provided hope that this was possible. The long-term objective was improved on-line quality control, with the side benefit of better process understanding. The analysis was restricted to measurements readily available on-line to enable future use of the procedure in on-line control.

Although all paper property measurements were investigated, this paper will focus on machine-direction tensile energy absorption (MD TEA), which was the key newsprint quality measurement at the study mill. Pulp quality measurements were available for each of the three main refiner lines, the rejects refiner line, the inlets to the two high-density storage tanks, and the machine chest. Important process measurements were also included in the analysis. The original analysis was completed in August 2009, but a new analysis was started in December 2010 to account for several major process changes.

ANALYSIS OF PAPER QUALITY

The first barrier to correlating with paper quality arose from the paper machine operation itself. Changes in TMP pulp quality would result in paper machine speed changes to keep the MD TEA at or above its target. This removed some of the variations coming from the TMP mill and hid the link between pulp quality and MD TEA. As a result, simple regression analysis did not work. A plot of paper machine reel speed versus MD TEA actually showed increased strength at higher speeds, which was the opposite of the expected trend.

This problem was solved by calculating short-term slopes (for four reels at a time) and using only data sets covering a speed change of at least 10 m/min. Data for the first reel after a shutdown were also excluded. Then outlier values were removed under the assumption that outliers indicated process upsets. When working with process data, one must remove outlier data, or they will tend to dominate the relationships. An operating mill is not a laboratory from which all disturbances can be excluded. Rather, one must assume that disturbances exist and compensate for them by using large quantities of data and systematically excluding questionable data.

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The remaining slopes were averaged, and the average was found to be statistically different from zero (Table 1). For clarity in graphical presentation, the slopes were sorted by paper machine speed, and 12-value averages were taken (Fig. 1).

The results for paper machine reel speed versus MD TEA are shown in Table 1. The data were collected from August 2009 through August 2010. This was a repeat analysis which was performed after a new shoe press was installed. Table 1 indicates that a speed increase of 100 m/min will typically drop the MD TEA by 0.59 J/m². Figure 1 indicates that the effect on MD TEA increases as the paper machine speed increases. This is consistent with observations by machine tenders and back-tenders. The working rule before installation of the shoe press was that a speed increase of 100 m/min typically reduced the MD TEA by 1.0 J/m².

### ANALYSIS OF TMP QUALITY

The next step was to build a model for speed-adjusted MD TEA based on pulp quality measurements. Pulp quality measurements were available after each of the three main refiner lines, after the rejects refiner, at the inlet of each high-density storage tank, and at the machine chest. Paper quality was measured for each reel produced. Key process measurements were included to enable screening of the data for downtime and outliers as well as for building models.

The data from the three main refiner lines were then combined. Six-hour averages were calculated for all variables to remove the effects of time delays in the storage tanks. All the measurements from the pulp quality analyzer were included. All the measured paper properties were included plus the speed-adjusted MD TEA.

Factor network analysis rather than regression analysis was used to create the model. Factor network analysis assumes that the input variables are correlated to some extent and represent a smaller set of truly independent variables. Process data usually fit this description. Although good models can be built using factor analysis, they cannot be treated in the same way as regression models. When using a regression model, it is possible to assume that all the variables are independent. The researcher can vary one independent variable and calculate a new value of the dependent variable. In a factor network model, all the variables change simultaneously. For instance, an increase in percent shives is accompanied by increases in freeness and fibre length. Therefore, one should always use a complete, real input data set when using factor network models.

At the beginning of the analysis, it was assumed that both main-line refiner and rejects-refiner measurements would be required; however, it was found that only the main-line refiner quality was needed to predict the speed-adjusted MD TEA. Adding the rejects refiner data improved the R-squared from 56% to 58%, but doubled the amount of data required. This probably occurred because the rejects system was production-limited and therefore the ability to adjust the final rejects quality was also limited. As a result, the rejects quality varied in a similar fashion to main-line quality. This result was also interesting because the measurements were made upstream of the main-line screening system, yet they predicted the final pulp quality as well as inlet samples from the main-line high-density storage tank.

Figure 2 shows how the speed-adjusted MD TEA captures both the measured strength and the effect of paper machine speed. The average paper machine speed was dropping gradually during May 2009, but the measured MD TEA remained relatively steady. Figure 3 compares the speed-adjusted MD TEA to the main-line refiner predicted MD TEA (R-squared = 56%). The effects of multiple refiner plate changes are also visible in both figures.

The prediction equation used was:

\[
\text{Pred. MD TEA} = -0.010 \times \text{(Freeness)} - 1.223 \times \text{(% Total Shives)} + 0.104 \times \text{(% Fibre Fraction 1)} + 0.183 \times \text{(% Fibre Fraction 4)} - 0.029 \times \text{(% Fibre Fraction 5)} + 0.799 \times \text{(% Fibre Fraction 6)} + 5.813
\]

### TABLE 1

<table>
<thead>
<tr>
<th>MDTEA versus Paper Machine Reel Speed (J/m² / m/min)</th>
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</thead>
<tbody>
<tr>
<td><strong>Average Slope</strong></td>
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<td>-0.0059</td>
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<td><strong>95% Confidence Limits</strong></td>
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<tr>
<td><strong>Upper</strong></td>
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<td>-0.0039</td>
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</table>

**Fig. 1 - Effect on MD TEA with Increased Paper Machine Speed.**

**Fig. 2 - Slope - MD TEA versus Paper Machine Real Speed.**

\[
y = -3.0E-07x^2 + 3.4E-04x - 2.2E-01
R^2 = 0.11
\]
Fibre fraction 1 represents the fines and fibre fraction 6 the longest fibres. The percentages of total shives and of fibre fraction 6 proved to be the most important variables in the equation.

A process model such as this one includes all the process factors measured by the input variables. For instance, changes in the paper machine white-water quality show up in the pulp quality measurements, but have nothing to do with refining. This should not be considered a weakness. Instead, it is recommended to watch for times when the results deviate unexpectedly and to use this information to find the source of unexpected and often unmeasured process disturbances. For instance, by watching the predicted MD TEA daily, one can observe how a change to one refiner line, such as replacing the refiner plates, affects the other refiner lines through the TMP white-water system.

After these relationships had been used successfully by the pulping superintendent for several months, in April 2010 they were added to the mill information system as calculated values so that anyone could track them on a day-to-day basis. This enabled personnel with varying levels of expertise to use them. For instance, the paper machine process engineer was able to spot the effect of a paper machine white-water disturbance on the MD TEA.

Each TMP refiner line had its own predicted TEA. This reduced the freeness, fibre length, and percent shives to a single number which could be used to evaluate pulp quality. Once a single number was available, it became quite apparent when a particular line had an upset or was in poor condition. For instance, the rotation of the refiners was reversed to extend refiner plate life. The MD TEA prediction showed poor pulp quality being produced for one rotation direction in August and September 2010, as shown in Fig. 4.

In addition to daily troubleshooting, the predictions for the main-line refiners were used to monitor refiner plate condition. For instance, the daily information plots shown in Fig. 5 reveal the results of a plate change on the Line 1 primary refiner. The bar graph quickly shows the status of each refiner line. The low predicted MD TEA on Line 1 prompted a primary refiner plate change, the impact of which could be seen in the following day’s trend.

As mentioned above, a second analysis is in progress at this writing, with the intention of taking major process changes into account. This analysis will also look at how to make better use of the information from the pulp quality analyzer. The new models have been able to produce similar results without using the freeness measurement. This is of particular interest in the main-line refiners because there the
freeness measurement is subject to latency effects.

**SUMMARY**

A predictive model for MD TEA on a newsprint machine has been developed from main-line refiner on-line pulp quality measurements. It has proved useful in process troubleshooting and provides a basis for determining when to change a set of refiner plates. It also has the potential for use in advanced quality control. Making this possible required both careful analysis and the improved fibre quality information provided by the pulp quality analyzer.

Compared to a decade ago, current pulp quality analyzers provide much improved fibre quality information. Both vendors and mills should look for opportunities to take advantage of this improvement. Combining the measurements into an average fibre length wastes some of the available information. The model discussed here has already used pulp quality to predict a paper strength property. Other attempts at modelling have indicated that the fibre size distribution could be used to predict freeness, or as a replacement for freeness. Some of these possible models will be mill-specific, but the possibility of identifying more fundamental relationships seems to exist.

This paper has also discussed methods beyond multiple linear regression for extracting information from process data. Short-term slopes can be analyzed in a spreadsheet program. Specialized software was required for the factor network analysis. These techniques are useful when one is unable to conduct a rigorous trial because of the cost and production considerations that always exist in a mill environment.

This paper represents a team effort. Discussions with various people in the industry have provided ideas and direction. My coworkers at Alberta Newsprint Company have also been supportive, and their observations and ideas have been essential.
Fig. 6 - Main-Line Predicted MD TEA Trend.

REFERENCES

HOW MICROSLUDGE® AND ANAEROBIC DIGESTION OR AEROBIC STABILIZATION OF WASTE ACTIVATED SLUDGE CAN SAVE ON SLUDGE MANAGEMENT COSTS

ABSTRACT

Waste activated sludge (WAS) is a waste by-product that is generated in large quantities in air-activated sludge or UNOX effluent treatment plants (ETPs) in pulp and paper mills. Currently, WAS is managed as a waste and is either burned (in spite of its high water content and negative fuel value), landfilled, or land-applied. However, WAS has both economic and environmental value: it is a potential source of both renewable energy and nutrients that can reduce the net costs of sludge management as well as a pulp mill’s environmental footprint.

Anaerobic digestion converts only a small portion of untreated WAS to form biogas, and the process is slow. Aerobic stabilization of raw WAS occurs in ETPs during long solids retention times, but results in high aeration costs. MicroSludge is a sludge pre-treatment process that ruptures the microbial cell membranes of WAS. This study shows that MicroSludge significantly increases the rate of anaerobic digestion of pulp and paper WAS, reduces the amount of waste sludge for disposal, and enables recovery of fertilizer that otherwise would need to be purchased to operate the ETP. This study also indicates that aerobic stabilization of MicroSludge-processed WAS results in rapid destruction of the sludge for disposal.

INTRODUCTION

MicroSludge uses a high-pressure homogenizer—a “cell disrupter”—to disrupt the microbial cells of waste activated sludge. (A cell disrupter consists of a positive displacement pump and a specialized cell-disruption valve.) By liquefying WAS cells, the rate-limiting step of hydrolysis is de-bottlenecked, speeding up WAS digestion. To boost sludge solubilization, a small amount (500 mg Na/L) of caustic (NaOH) solution is added to WAS before cell disruption. Figure 1 shows the cell-disruption valve.

The scanning electron micrographs in Fig. 2 show the effect of MicroSludge on WAS. The cytoplasm of the microbial cells in the WAS is released when the cells are lysed, leaving only remnants of the cell membranes in particulate solid form. The cytoplasm, once released from the cells, is rich in bio-available degradable carbon and nutrients.

There are two options for integrating MicroSludge into an existing ETP at a pulp and paper mill:

Option 1 - Anaerobic Digestion: WAS from the secondary clarifier is thickened, MicroSludge-processed, anaerobically digested, and de-watered along with primary sludge. The de-watering liquid is returned...
to the activated-sludge ETP. Figure 3 shows how an ETP can be modified to include both MicroSludge and anaerobic digestion.

By liquefying WAS, MicroSludge enables anaerobic digesters to convert the anaerobically biodegradable portion of WAS rapidly into biogas (primarily methane and carbon dioxide), reducing the amount of WAS going to de-watering and disposal. Digester effluent is de-watered, and the relatively low-BOD, but nitrogen- and phosphorus-rich, de-watering liquid is returned to the activated-sludge ETP. This approach produces biogas that can offset fuel purchases, reduce the need for polymer for de-watering, and displace a portion of the nitrogen- and phosphorus-fertilizers that are needed to treat the carbon-rich mill effluent.

**Option 2 - Aerobic Stabilization**: WAS from the secondary clarifier is thickened, MicroSludge-processed, and de-watered along with primary sludge. (No anaerobic digester is used.) The de-watering liquid is returned to the activated-sludge ETP. Figure 4 shows how an ETP can be modified to include the MicroSludge process without also including an anaerobic digester.

By returning de-watered MicroSludge-processed WAS directly to the activated-sludge ETP without anaerobic digestion, the cost of building and operating an anaerobic digester is avoided, although no biogas is produced. Aerobic stabilization of MicroSludge-processed WAS rapidly reduces the amount of WAS going to de-watering and disposal and returns the nitrogen- and phosphorus-rich de-watering liquid back to the activated-sludge ETP. This reduces polymer consumption and displaces some of the nitrogen and phosphorus fertilizers used in the ETP.

By including a sludge management solution, ETPs can re-consider their practice of operating with a long sludge age. This practice minimizes WAS production, fertilizer requirements, use of polymer for de-watering, and the amount of sludge requiring disposal, but at the cost of high aeration requirements. By liquefying the sludge, less of the sludge will require polymer addition and disposal, and the nitrogen and phosphorus in the WAS can be recycled, enabling operation at a lower sludge age, increased (20% to 25% greater) aeration efficiency [1], and therefore lower electrical costs.

**METHODS**

Anaerobic Digestion: Two laboratory-scale anaerobic digesters were used: a batch biogas potential test apparatus and a set of daily-fed digesters. The batch system used bottles in an incubator/shaker maintained at 35°C with biogas volumes and concentrations measured daily. Liquid assays (TS, VS, COD, TKN, NH₃, and PO₄) were performed at the end of each run. In addition, a set of semi-continuous digesters made up of two 10-litre, continuously mixed 35°C digesters were fed once daily with raw WAS and MicroSludge-processed TWAS until near-steady-state performance was measured at each of the hydraulic retention time (HRT) settings.

Aerobic Stabilization: 20-litre, continuously aerated (to maintain at least 2.0 mg/L dissolved oxygen concentration) activated sludge containers were batch-fed and maintained at 20°C (room temperature). Particulate solids destruction and

---

**Fig. 1 - Cell-Disruption Valve.**

**Fig. 2 - Scanning Electron Micrographs: Raw WAS (left); MicroSludge-Processed WAS (right).**
sludge settle-ability testing in side-by-side testing of raw WAS and of 25% Micro-Sludge-processed WAS were the measures of performance used.

RESULTS

Technical: WAS Solubilization
Because micro-organisms consume liquids and not solids, liquefying WAS addresses the rate-limiting step of hydrolysis for either anaerobic digestion or aerobic stabilization. For the two samples of pulp mill WAS tested, MicroSludge increased the scOD/tCOD of the WAS by 5.9 to 10.7 times, as summarized in Table 1. MicroSludge reduced the viscosity of WAS by 95%.

Anaerobic Digestion
The objective of anaerobic digestion is to produce biogas for fuel, reduce the amount of solids going to de-watering and disposal, and return the nitrogen and phosphorus in the de-watering liquid back to the activated-sludge ETP to displace purchased fertilizers. Figures 5 and 6 and Table 2 illustrate the effect of MicroSludge on each of these parameters.

Figure 5 shows the effect of MicroSludge compared to raw WAS on biogas production and composition for anaerobic digester HRTs ranging from 3 to 20 days. The relative increase in biogas produced per mass TSS of WAS fed ranged from 18% greater at the longest HRT to almost five times more biogas at the shortest HRT. For all the HRTs tested, the average methane content of the biogas was 12.4% greater than for raw WAS, corresponding to increased energy content in the gas.

Eskicioğlu and Saha [2] tested MicroSludge on batch anaerobic digestion of WAS from a CTMP mill. They observed increases in specific methane yields (mL/mg TCOD added) of 60% for digestion at 35°C and 72% for digestion at 55°C compared to the control digester after 21 days in the MicroSludge digester at STP (1 atm, 0°C). After 42 days of batch digestion, 34% and 16% more biomethane was produced by the MicroSludge-fed 35°C and 55°C digesters.

Figure 6 shows the effect of MicroSludge compared to raw WAS on reduction of solids for de-watering and disposal. MicroSludge plus anaerobic digestion averaged a 48.4% reduction in particulate solids, compared to 23.4% for raw WAS plus digestion. For all the HRTs tested, particulate solids reduction was far greater with MicroSludge, which corresponds to less polymer required for de-watering and less sludge going to disposal.

Table 2 summarizes the effect of MicroSludge compared to raw WAS on the concentrations of soluble COD, ammonia and phosphate in the de-watering liquid following anaerobic digestion.

For the range of digester HRTs tested, Table 2 shows that:
- TSS was reduced by an average of 23% by anaerobic digestion of raw WAS and by 49% by MicroSludge and anaerobic digestion. The small incremental
reduction of TSS or VSS achieved by anaerobic digestion of MicroSludge-processed WAS compared to MicroSludge processing alone indicates that the particulate solids following MicroSludge processing are only slightly degradable and can be removed by de-watering before digestion without compromising either biogas production or solids reduction.

- MicroSludge resulted in elevated sCOD levels in the de-watering liquid, most of which was degraded. Non-degradable sCOD does not add to the aeration load of the ETP.

- Anaerobic digestion of MicroSludge-processed WAS increased the ammonia concentration from 47 to 1,177 mg/L and the phosphate concentration from 18 to 279 mg/L. This would decrease the need for the ETP to purchase fertilizer to treat the mill effluent.

### Aerobic Stabilization

The objective of aerobic stabilization is to reduce the amount of solids going to de-watering and disposal and to return the nitrogen and phosphorus in the de-watering liquid back to the activated-sludge unit to displace purchased fertilizers without negatively impacting ETP operations, the secondary clarifier in particular.

Figure 7 shows the effect of MicroSludge compared to raw WAS on reduction of solids going to de-watering and disposal, using aerobic stabilization of WAS samples from a pulp and paper mill as well as from a municipal WWTP after 24 hours of batch digestion at room temperature.

For these two samples, as shown in Fig. 7, MicroSludge plus aerobic stabilization averaged a 46% reduction in particulate solids, compared to 21% for raw WAS plus aerobic stabilization.

No change in the settling rates of effluent from the activated sludge unit was measured (data not shown), indicating that MicroSludge processing is compatible with the operations of a secondary clarifier.

This batch testing will be followed up in future research with continuously
fed activated sludge to obtain a better understanding of the rates and limits of aerobic solids destruction using MicroSludge.

**Benefits for a Typical Pulp and Paper Mill**

The operating savings and capital costs of MicroSludge plus anaerobic digestion or aerobic stabilization will vary depending on the mill. Mahmood and Elliot [3] of FPInnovations generated preliminary estimates of economic benefits for “typical” mechanical and Kraft pulp mills in Canada. Operating benefits include the following:

- Generate clean, renewable energy from biogas to displace non-renewable fossil fuels to generate energy
- Recycle nitrogen and phosphorus from WAS to reduce fertilizer costs for effluent treatment
- Reduce the amount of WAS going to disposal, which also reduces polymer costs
- Reduce ETP electricity costs by increasing aeration efficiency by operating at a younger sludge age
- Reduce greenhouse gas (GHG) emissions.

As the value of renewable energy, fuel, electricity, fertilizers, sludge disposal, polymer, and GHGs increases, savings in operating costs will also increase.

**CONCLUSIONS**

1. MicroSludge and anaerobic digestion can reduce costs for sludge management in the following ways:
   - Less WAS will remain for de-watering and disposal, reducing polymer consumption for de-watering and disposal costs.
   - Less nitrogen and phosphorus fertilizer will be needed by the ETP.
   - Methane-rich biogas will be produced as a renewable fuel source.
   - Fast anaerobic digestion will decrease the size of the digesters needed, significantly lowering capital costs compared to conventional anaerobic digestion.  
2. MicroSludge and aerobic stabilization in an existing ETP can reduce costs and provide the following benefits:
   - Less WAS will remain for de-watering and disposal, reducing polymer consumption for de-watering.
   - Less nitrogen and phosphorus fertilizer will need to be added to the ETP.

**NEXT STEPS**

A transportable demonstration plant to validate the performance of MicroSludge with anaerobic digestion, aerobic stabilization, or both will shortly be operational at Catalyst Paper's pulp and paper mill in Crofton, British Columbia. Significant improvements in the design and operation of both the anaerobic digesters and the activated sludge units are expected to result in increases in biogas production, solids reduction, and soluble nitrogen and phosphorus in the de-watering liquid.

Figure 8 shows the MicroSludge and anaerobic digester module. The plant also includes: (a) a sludge thickening module, housing both a rotary drum thickener and centrifuge; (b) a module containing two parallel activated-sludge units; and (c) a module containing an analytical laboratory, laboratory-scale digesters, and an office.

A 15-month testing program will be used to determine how to optimize performance and minimize capital costs for permanent operations. The demonstration plant will be used for future trials at other pulp mills.
REFERENCES


Fig. 8 - MicroSludge and Anaerobic Digestion Demonstration Plant.

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ABSTRACT

This paper presents a model of Kraft char bed burning which considers gas-phase chemical reactions inside the boundary layer. The model was used to simulate the Institute of Paper Science and Technology char-burning experiments to study the effect of gas-phase reactions on char burning.

The simulations showed that during combined oxidation and gasification, oxygen is consumed inside the boundary layer. Oxygen can be consumed to such an extent that carbon conversion involving gas-phase reactants takes place entirely through gasification reactions. In general, the effect of the gas-phase reactions is to decrease overall burning rate and increase bed temperature. Factors determining the magnitude of the decrease in burning rate are char gasification reactivity and boundary-layer mass transfer. The model predictions were found to be in reasonable quantitative agreement with experimental burning data on carbon conversion, bed temperatures, and carbon oxide formation during combined oxidation and gasification.

INTRODUCTION

Boundary-layer mass transfer is important in recovery boiler char-bed burning. Oxidation of carbon in char by oxygen, carbon dioxide, and water vapor is considered to be limited to a great extent by their gas-side mass transfer rates [1]. Increased insight into the processes occurring inside the boundary layer is important for a deeper understanding of char-bed burning. This knowledge is also important in the development of furnace models based on computational fluid dynamics (CFD).

The term reactive boundary layer describes the situation in which gas-phase reactions take place inside the boundary layer [2]. The current understanding of reactive boundary layers in black liquor char-bed burning is based on experimental work carried out at the Institute of Paper Science and Technology (IPST) [3–5]. This IPST research generated valuable experimental data and provided a first insight into the phenomenon. The IPST experiments are described in more detail in a separate section below.

Reactive boundary layers have also been considered in black liquor combustion modelling. For example, Wåg et al. [6] and Sutinen et al. [7] used a simplified model to consider the effect of the reactive boundary layer on overall char-burning rate. Wåg et al. modelled combustion of a single black liquor droplet or particle, whereas Sutinen et al. carried out simulations of the IPST char-bed burning experiments. The predictions of the Sutinen et al. model showed relatively good agreement with experimental data for burning rates and bed temperatures. The approach used by Wåg et al. and Sutinen et al. followed the “mutually exclusive” concept proposed by Grace et al., which states that during simultaneous oxidation and gasification, the overall burning rate is given by the greater of two quantities: the carbon conversion by oxygen and the sum of the conversions by gasification reactions [3]. In this simplified model, gas-phase reactions are not explicitly considered; the conversions by each gas-phase reactant are calculated assuming a non-reactive boundary layer.
From a computational perspective, a simplified model such as that proposed by Grace et al. has the advantage that there is no need to resolve the boundary layer in detail.

Other models have included more detailed descriptions of reactive boundary layers. For example, Järvinen used a two-film model to study the effect of the envelope flame formed around a black liquor droplet during overlapping pyrolysis and char conversion [8]. Ip et al. developed a one-dimensional droplet model in which the calculation grid could be extended into the gas phase to consider gas-phase reactions [9]. Both Järvinen and Ip et al. found that the reactive boundary layer can suppress the oxygen flux to the char surface and that char conversion then occurs mainly by gasification reactions involving carbon dioxide and water vapor. Although these studies provided additional insight into the phenomenon, it can be argued that the effect was discussed only on a general level—a systematic study of relevant parameters was not carried out.

To achieve a better understanding of reactive boundary layers, some insight could possibly be gained from studies of fuels other than black liquor. Reactive boundary layers around coal-char particles have been analyzed, for example, in textbooks by Law [2] and Glassman [10]. Law considered carbon reactions with oxygen (C + ½O₂ → CO) and carbon dioxide (C + CO₂ → 2CO) as well as a gas-phase reaction (½O₂ + CO → CO₂). Law found that the impact of the gas-phase reaction on (i) the amount of oxygen reaching the carbon surface, (ii) the extent to which carbon conversion occurs by oxygen and carbon dioxide, and (iii) how the overall carbon conversion is affected depends on the gasification reactivity of the char. The analysis performed by Law indicated that char gasification reactivity could also be an important parameter in Kraft char burning in the presence of a reactive boundary layer. However, Law did not consider carbon conversion by water vapor or by sulphate.

It appears that there is a need for increased understanding of reactive boundary layers in black liquor char-bed burning. This paper aims to provide a better understanding by means of mathematical modelling. First, a short summary of the IPST experiments is given, followed by a model description. The model is then used to simulate the IPST experiments to study the effects of reactive boundary layers on char-bed burning. Model predictions are also compared with the IPST experimental results.

**IPST EXPERIMENTS**

In the IPST experiments, black liquor char was burned in a laboratory reactor. Burning was carried out using a gas jet sweeping over the char bed. The bed rested on a tray which was moved upwards to sustain the burning geometry. Some of the experiments were aimed specifically at studying char burning when using more than one oxidant. Three oxidants in nitrogen—oxygen, carbon dioxide, and water vapor—were used in four combinations. Table I summarizes the inlet gas compositions and the char-burning results in the IPST experiments.

In the IPST experiments, the overall oxygen content of the gas jet increased from Case 1 to Case 4. One of the findings was that overall carbon conversion did not increase with more available overall oxygen. Including water vapor in the gas feed (Cases 3 and 4) resulted in a decrease in carbon conversion and an increase in bed temperature. In addition, when water vapor was present in the gas feed, practically all carbon oxides exiting the reactor were in the form of carbon dioxide. The differences in carbon conversion, bed temperature, and exhaust gas composition were attributed to gas-phase reactions in the boundary layer. The water vapor in the gas jet catalyzed the oxidation of carbon monoxide to carbon dioxide.

### TABLE 1

<table>
<thead>
<tr>
<th>CASE</th>
<th>O₂ (%)</th>
<th>CO₂ (%)</th>
<th>H₂O (%)</th>
<th>Conversion (mmol/s-m²)</th>
<th>Temperature (°C)</th>
<th>CO/(CO+CO₂)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Case 1</td>
<td>14</td>
<td>-</td>
<td>-</td>
<td>243.5</td>
<td>961</td>
<td>0.504</td>
</tr>
<tr>
<td>Case 2</td>
<td>14</td>
<td>10</td>
<td>-</td>
<td>324.5</td>
<td>932</td>
<td>0.638</td>
</tr>
<tr>
<td>Case 3</td>
<td>14</td>
<td>-</td>
<td>10</td>
<td>248.6</td>
<td>990</td>
<td>0.011</td>
</tr>
<tr>
<td>Case 4</td>
<td>14</td>
<td>10</td>
<td>10</td>
<td>280.9</td>
<td>967</td>
<td>0.000</td>
</tr>
</tbody>
</table>
MODEL DESCRIPTION

The model structure is illustrated in Fig. 1. The Fluent CFD software [11] formed the basis of the model. The boundary layer was resolved from the free stream to the char-bed surface. User-implemented sub-models included a modification to the standard turbulence-chemistry interaction model and a char-bed burning model.

Reactions

A two-step gas-phase reaction mechanism (R1-R2) was used, based on the mechanism proposed by Jones and Lindstedt [12]. Char-carbon reactions included direct oxidation (R3), gasification reactions with carbon dioxide (R4) and with water vapor (R5), and sulphate reduction (R6).

Gas-phase reactions

\[
\begin{align*}
H_2 + \frac{1}{2}O_2 & \rightarrow H_2O \\
CO + H_2O & \leftrightarrow CO_2 + H_2 \\
\end{align*}
\]

Char-carbon reactions

\[
\begin{align*}
C(s) + \frac{1}{2}O_2(g) & \rightarrow CO(g) \\
C(s) + CO_2(g) & \rightarrow 2CO(g) \\
C(s) + H_2O(g) & \rightarrow CO(g) + H_2(g) \\
C(s) + \frac{1}{2}Na_2SO_4(l) & \rightarrow CO_2(g) + \frac{1}{2}Na_S \\
\end{align*}
\]

Gas-phase and mass transfer

The Eddy break-up model available in the CFD software describes turbulence-chemistry interactions in the gas phase. For each reaction, it considers (i) the rate of chemical kinetics, as calculated using equations of the Arrhenius form and time-averaged species concentrations, and (ii) the rate of turbulent dissipation, i.e., the rate of mixing on a molecular level. The effective rate of a reaction is calculated as the lesser of the kinetic rate and the turbulent mixing rate. The standard model is valid in the fully turbulent region of a turbulent flow, i.e., far from solid surfaces [13].

In this work, the standard turbulence-chemistry interaction model has been modified. Equation 1 gives the modified expression for turbulent mixing rate. Turbulent and molecular viscosities are used to calculate a weighted average of the original turbulent mixing rate and the kinetic rate. As result of this modification, near a solid surface, the effective rate of reaction (Eq. 2) is determined by chemical kinetics. The effect of the modification diminishes smoothly with distance from the surface, and outside the boundary layer, the effective reaction rate is determined by the standard turbulence-chemistry interaction model.

\[
R_{\text{mix,mod}} = \frac{H_{\text{t}} \cdot R_{\text{mix}} + H_{\mu} \cdot R_{\text{tuf}}}{H_{\mu} + H_{\mu}}
\]

\[
R_{\text{eff}} = \min\left(R_{\text{mix,mod}}, R_{\text{tuf}}\right)
\]

Gas-phase diffusion was modelled as multi-component diffusion using the Maxwell-Stefan equations [11]. The binary diffusion coefficients between each pair of gas species were calculated based on the kinetic theory of gases and using Lennard-Jones potentials [14].

Char-bed chemistry and mass transfer

Direct oxidation of char carbon (R3) was assumed to be infinitely fast and was modelled as mass-transfer limited. The oxygen flux to the bed surface was calculated using Eq. 3. The oxygen-concentration gradient was calculated from the oxygen concentration in the next-to-bed cell assuming zero concentration at the bed surface. Diffusion to the bed surface was assumed to be Fickian, an assumption supported by the fact that at low concentrations of the diffusing species, the Maxwell-Stefan equations reduce to Fick’s law of diffusion [15]. The diffusion coefficient of oxygen in the gas mixture was calculated from the binary diffusion coefficients using Eq. 4. The reaction rate per unit of char-bed surface area (mol/s-m²) was calculated from the oxygen flux assuming the reaction stoichiometry (R3).

\[
N_{O_2} = D_{O_2,\text{gas}} \frac{\Delta C_{O_2}}{\Delta x}
\]

\[
D_{O_2,\text{gas}} = \frac{1 - y_i}{\sum_{j \neq i} (y_j / D_{t,j})}
\]

Gasification reactions (R4-R5) were modelled as having a finite rate. The kinetic expressions were based on the work by Li and van Heiningen [16,17] and are stated as Eqs. 5 and 6.

\[
R_{C-CO_2} = 3.94 \times 10^9 A_{C} C_{CO_2} e^{-29000 / (T + 273.15)} \frac{C_{CO_2}}{C_{CO_2} + 3.4 \cdot C_{CO}}
\]

\[
R_{C-H_2O} = 7.1 \times 10^8 A_{C} C_{H_2O} e^{-21000 / (T + 273.15)} \frac{C_{H_2O}}{C_{H_2O} + 1.42 \cdot C_{H_2}}
\]

The specific surface area ($A_s=160$ m²/g) was assumed to be equal to that in the experimental char investigated by Li and van Heiningen [16]. The char-bed carbon concentration ($C_C=220$ mol/m³) was calculated from the reported bulk density and the carbon mass fraction of the IPST char bed [4].

The gasification reactions were modelled following the principles of first-order kinetics and pore diffusion (Eqs. 4 and 7–11) [18]. A first-order apparent intrinsic reaction rate was calculated using species concentrations in the next-to-bed cells. For the calculation of pore diffusion (Eq. 8), the tortuosity factor and the porosity of the experimental char bed were not known. Therefore, the tortuosity factor and the char porosity were modelled as a lumped parameter (net tortuosity factor). As a first assumption, a net tortuosity factor of 2 was used, which means that the effective diffusivities of carbon dioxide and water vapor inside the char bed were assumed to be half their respective values in the gas mixture above the char bed. The net tortuosity factor of 2 corresponds to pores at 45° angles to the nominal direction into the bed, an assumption used in the droplet model proposed by Järvinen [8]. In contrast, based on the relatively low bulk density of the experimental char bed, Sutinen et al. assumed that the diffusivity
inside the bed was equal to the diffusivity in the gas phase [7]. This corresponds to a net tortuosity factor of 1. The assumption of Sutinen et al. was tested in simulations of IPST cases 1–4. The net tortuosity factor did not have a significant impact on the predictions. Decreasing net tortuosity from 2 to 1 increased burning rates and decreased bed temperatures by only a few per cent.

\[
\frac{r}{C_X} = \frac{X_{\text{bed surface}}}{C_X}
\]  

(7)

\[
D_{\text{eff}} = \frac{D_{\text{gas}} \cdot \epsilon}{\tau}
\]  

(8)

\[
\phi = L \cdot \sqrt{\frac{r}{D_{\text{eff}}}}
\]  

(9)

\[
\eta = \frac{\tanh(\phi)}{\phi}
\]  

(10)

\[
\frac{R_{C-X}}{A} = \eta \cdot r \cdot L \cdot C_{X,\text{bed surface}}
\]  

(11)

The active burning layer of the experimental char bed has been reported to have a thickness of 0.03–0.07 metres [4]. A value of 0.03 metres was used here, which determines the maximum distance into the char bed over which gasification reactions can take place. The actual distance over which reactions occur depends on the intrinsic reactivity. For example, in simulations carried out for the purpose of comparing predictions to experimental data, the reactions took place over a distance of a few millimeters. The gasification reaction rates per unit of char-bed surface area (mol/s-m²) were calculated using Eq. 11, which can be considered to be composed of a kinetic rate coefficient \(k = \eta r L\) with units (m/s) and a reactant concentration at the bed surface \(C_{X,\text{bed surface}}\) with units (mol/m³). When presenting results, the ratio between the external mass-transfer coefficient and the kinetic rate coefficient was used to characterize the gasification reactivity of the char.

Although the gasification reactions were modelled as having a finite rate, the case of mass-transfer-limited conversion can be simulated by setting the intrinsic rates of the gasification reactions to high values.

**Sulphate reduction** was modelled by considering reaction kinetics and the amount of sulphate present in the char. Re-oxidation of sulphide was not considered. The kinetic rate of sulphate reduction was calculated using Eq. 12 [6]. Values for moles of carbon and sulphate per mole Na₂ in char were estimated as 2.5 and 0.1 respectively based on reported char properties [4]. Sodium in char was calculated as 635 moles Na₂/kg char, and a char bed bulk density of 14.2 kg char/m³ was assumed [4]. Carbon conversion kinetics by sulphate reduction per unit of surface area were calculated using Eq. 13 in combination with the reaction stoichiometry (R6). Equation 12 gives the instantaneous rate of sulphate reduction. In a char particle, this rate would change with time as the sulphate was depleted. This time dependence is not included in the model.

Based on the reported char composition [4], it was estimated that there was enough sulphate in the char to convert 7.9% of total organic carbon. For additional sulphate-containing char to enter the burning zone, the remaining carbon must be converted by the reactants from the gas jet. Therefore, in the model, the rate of carbon conversion by sulphate was limited to 7.9/100-7.9) = 8.6% of the rate of conversion by gas-phase reactants. Below this limit, the rate of carbon conversion by sulphate was determined by chemical kinetics.

\[
R_{C-SO_4}^{\text{kin}} = \left(3790[C][SO_4]^1 \cdot e^{-\frac{78100}{RT}}\right)Na_{2-\text{m-char}} \cdot \rho_{\text{charbed}}
\]  

(12)

\[
R_{C-SO_4}^{\text{kin}} / A = R_{C-SO_4}^{\text{kin}} \cdot L
\]  

(13)

**Char-bed mass and energy balance**

The bed surface temperature was calculated using a mass and energy balance. Figure 2 illustrates the balance terms. Carbon conversion was calculated for each position along the char bed, whereas an average bed temperature was calculated for the whole bed length. The bed temperature and burning rate are interconnected and were calculated iteratively as part of the overall CFD solution.

The char-bed energy balance considers heat from char carbon reactions; the enthalpy needed to raise the char temperature from its pre-heating value to burning temperature; the enthalpy of gases entering and leaving the bed; heat transfer by conduction between the bed and the gas phase; and the enthalpy of the smelt removed from the top of the char bed. Carbon conversion, as well as gases entering and leaving the char bed, was calculated as the sum of the individual values in each location along the length of the bed. For heat conduction, an average conductive heat-transfer coefficient and an average gas temperature above the bed were used. Inorganic salts were assumed to be in the liquid phase only; transitions to or from the solid state were not considered.

For the heats of the char-carbon reactions, values of -113, 169, 110, and 109 kJ/mole carbon were used for reactions (R3-R6) respectively [19]. The rate of pre-heated char entering the balance was determined from the predicted carbon conversion by considering the fraction of convertible carbon in char. This amount was estimated from the reported char composition as 0.2 g carbon/g char by assuming total elemental carbon to be split into convertible carbon and carbonate [4]. The pre-heating temperature of the char was assumed to be 775°C based on reported values (750°C–800°C) [4], and the
specific heat capacity of the char was assumed to be 2 kJ/kg-K \[1\]. The specific heat capacities of the gases entering and leaving the bed were assumed equal and were obtained from the CFD solver.

A conductive heat transfer coefficient at the bed surface was determined from the heat conductivity of the gas and the distance from the bed surface to the center of the next-to-bed cell. The amount of smelt leaving the balance per unit of converted carbon was estimated as 4 grams smelt/gram carbon from the reported char composition \[4\], and the specific heat capacity of the smelt was assumed equal to that used for the char.

The gas-phase chemistry mechanism used in this work was developed for situations where water vapor is present. In the IPST experimental Cases 1 and 2, there was no water vapor in the gas jet. To test the model in these cases also using the gas chemistry mechanism, hydrogen was assumed to be released from the char bed. The release of hydrogen was assumed to occur in proportion to the amount of carbon converted. From the reported elemental composition of the experimental char \[4\], a hydrogen release ratio of 0.18 moles H$_2$/mole C was used.

**RESULTS AND DISCUSSION**

In this section, the results of the simulations of the IPST experiments are presented and discussed. First, observations on the impact of gas-phase reactions on species concentrations, temperatures, and velocities inside the boundary layer are presented. Second, the impact on char-burning rate is discussed. Finally, model predictions are compared against the IPST experimental results.

**Species, temperature, and velocity profiles**

Figure 4 presents examples of velocity, oxygen, and temperature contours from simulation of IPST Case 4. The boundary layer increases in thickness with distance downstream. Therefore, profiles of predicted variables depend on the downstream position. In general, the shapes of the profiles as a function of distance from the char bed are similar, but their thickness varies. Figures 5–8 present profiles taken at a position 5 cm downstream from the leading edge of the char bed. This position is also indicated in Fig. 4.

Figure 5 presents the predicted species mole fractions from IPST Case 1 simulations. Oxygen was present in the bulk gas as the only oxidant. In the non-reactive boundary-layer scenario, oxygen diffused into the char bed, and carbon monoxide, carbon dioxide, and hydrogen were released as char-carbon conversion products. When the boundary layer was chemically reactive, oxygen was consumed inside the boundary layer, and practically no oxygen reached the char surface. At the same time, the carbon dioxide and water vapor concentrations increased inside the boundary layer. As a result of gas-phase reactions, carbon conversion took place through all three reaction pathways (R3–R5) and mainly by gasification reactions, even though only oxygen was fed to the reactor. As will be shown, the release of hydrogen from the char bed resulted in an overprediction in Case 1 compared to experimental data of the extent to which gas-phase reactions occurred inside the boundary layer. However, a very low moisture content in the bulk gas can be expected to result in a similar situation, as illustrated in Fig. 5.

Figure 6 presents the predicted species mole fractions from simulation of IPST Case 4. In this case, all three oxidants were present in the reactor gas feed. Moreover, with this bulk gas composition, oxygen-consuming gas-phase reactions...
resulted in a decrease in carbon conversion by oxygen (R3) and an increase in conversion by gasification reactions (R4–R5). In Fig. 6, these changes are indicated by changes in mole fraction gradients at the bed surface.

Figure 7 presents temperature profiles from simulation of IPST Case 4. As a result of gas-phase reactions, there is typically a temperature maximum inside the boundary layer. The gas-phase reactions had a net exothermic impact on overall combustion. Because of heat transfer from the gas phase to the char bed, the bed temperature also remained high, so that carbon conversion occurred mainly through endothermic gasification reactions (R4–R5). The temperature peak coincides with a relatively narrow reaction front inside the boundary layer. The concentration profiles of oxygen and carbon monoxide overlap very little, i.e., oxygen does not co-exist with hydrogen or carbon monoxide (Fig. 6).

The gas flow inside the boundary layer was also affected by the increase in gas temperature. Figure 8 presents profiles of flow-direction and transverse boundary-layer velocities from simulation of IPST Case 4. Based on the flow-direction velocities, the boundary layer is approximately 20% thicker when it is chemically reactive. The transverse velocity was directed from the char surface towards the free stream. Figure 8 shows that the transverse velocity was greater in the case of a reactive boundary layer, particularly in the outer part of the boundary layer. This region extended approximately from the location of the temperature maximum towards the free stream. With increasing temperature, the gas density decreased, and the transverse velocity increased. Although not shown here, the diffusion velocities of oxygen, carbon dioxide, and water...
vapor in the outer part of the boundary layer were of similar magnitude to the transverse velocities. As a result, the increase in transverse velocity effectively decreased the net mass transfer rate from the bulk gas towards the char bed.

**Char-burning rate during combined oxidation and gasification**

Simulations of IPST Case 4 were carried out to study the impact of gas-phase reactions on char-burning rate during combined oxidation and gasification. In these simulations, the bed temperature was fixed at the experimental value of 967°C, and char gasification reactivity was varied from zero reactivity to external mass transfer-limited gasification. In practice, Eqs. 5 and 6 are multiplied by factors ranging from zero to 1000. The simulations presented here focussed on carbon conversion involving gas-phase reactants. Carbon conversion by sulphates was excluded.

The kinetic rate of gasification by carbon dioxide (R4) is generally lower than the kinetic rate of gasification by water vapor (R5). When the char gasification reactivity is increased by multiplying Eqs. 5 and 6 by the same factor, the reaction involving water vapor becomes mass transfer-limited before the reaction involving carbon dioxide does. Therefore, gasification by carbon dioxide is limiting for char-bed burning to become completely mass transfer-limited. In Figs. 9–13, carbon conversion is plotted as a function of $k/m$, where $k$ is the kinetic rate coefficient of the C-CO$_2$ reaction and $m$ is the external mass transfer coefficient characteristic of the simulated bed-burning system. A high value of $k/m$ indicates a char reaction limited by gas-side mass transfer. The kinetic rate coefficient for the C-CO$_2$ reaction was obtained from each simulation case by varying char-gasification reactivity. The mass-transfer coefficient was determined from the simulated case of carbon conversion by oxygen only with a non-reactive boundary layer.

In Fig. 9, when the gasification reactivity of the char is zero, conversion occurs due to reaction with oxygen only. As gasification reactivity increases, conversion by gasification reactions increases. There is a slight decrease in conversion by oxygen due to the increased outward flux of conversion products, but the overall conversion increases until it becomes mass transfer-limited by all reactions.

Figure 10 presents simulated carbon conversions in IPST Case 4, but now the boundary layer is reactive. At zero char-gasification reactivity, carbon conversion occurs by oxygen only. The conversion...
**Fig. 9** - Carbon conversions in the non-reactive boundary-layer case as a function of increasing char gasification reactivity.

**Fig. 10** - Carbon conversions in the reactive boundary-layer case as a function of increasing char gasification reactivity.

**Fig. 11** - Overall carbon conversion in the non-reactive and reactive boundary-layer cases as a function of increasing char gasification reactivity.

**Fig. 12** - Overall carbon conversion as a function of increasing char gasification reactivity with constant gas temperature. The small graphs show the temperature profile as a function of distance from the bed surface, which indicates the temperature field used in the various calculations.

**Fig. 13** - Illustration of carbon conversion according to the mutually exclusive concept in the case of decreased oxygen and increased carbon dioxide and water vapor in the bulk gas.
is less than in the non-reactive situation because part of the oxygen reacts inside the boundary layer. With increasing gasification reactivity, conversion by oxygen decreases. At sufficiently high gasification reactivity, conversion is practically due to gasification reactions only. The share of carbon conversion by gasification reactions R4 and R5 varies as a function of char gasification reactivity, which is affected by the water-gas-shift reaction (R2) at the char surface.

Figure 11 presents a comparison between the overall carbon conversion curves shown in Figs. 9 and 10. In general, the conversion curves are similar in shape. However, conversion is consistently lower in the reactive boundary-layer situation than in the non-reactive one: approximately 40% lower at zero gasification reactivity and approximately 10% lower at high gasification reactivity.

The decrease in overall conversion seen in Fig. 11 results from both the direct effect of boundary-layer chemistry and the decrease in the boundary-layer mass transfer due to the higher transverse gas velocity. To elucidate further the impact of boundary-layer chemistry, additional simulations were carried out. Starting from the non-reactive boundary-layer case, the gas-phase reactions were allowed to affect species concentrations, but not gas temperature. A corresponding calculation was also carried out using a gas-temperature field with a temperature maximum inside the boundary layer. The decoupling of gas temperature and chemical reactions is physically unrealistic, but serves the purpose of studying what happens when all predicted changes in carbon conversion are due to boundary-layer chemistry only. Figure 12 presents the results of these simulations.

In Fig. 12, regardless of the temperature field used, there is no decrease in overall conversion when char-carbon reactions occur at a mass transfer-limited rate. The decrease in carbon conversion by oxygen is fully compensated for by the increase in carbon conversion by gasification reactions. These results agree with those reported by Law, who assumed a constant mass-transfer coefficient [2]. These results are significant for possible future model simplifications. The implication is that, for a given boundary-layer mass-transfer coefficient, the overall mass transfer-limited burning rate in the presence of gas-phase reactions can be calculated without having to resolve the species profiles inside the boundary layer.

Figure 12 also shows that when char gasification reactivity is low, overall carbon conversion is decreased. At zero gasification, the decrease depends on the gas temperature field used. The decrease is approximately 40% when using the temperature field with a temperature maximum, but only approximately 30% when using the temperature field without the maximum. This result is also consistent with those presented by Law, who found that in the limiting case of an infinitely fast gas-phase reaction and no gasification, conversion was decreased by 50%. Under these assumptions, half the oxygen diffusing towards the char surface reacts with carbon, and the other half reacts with carbon monoxide in the gas phase. In the simulations reported here, the gas-phase reactions proceeded at a finite rate, resulting in more oxygen reaching the char surface than in the limiting case described by Law.

Figures 11 and 12 also show a comparison between the overall burning rate predicted by the proposed model and the rate calculated using the mutually exclusive concept proposed by Grace et al., in which the overall burning rate was given by the greater of the carbon conversion by oxygen and the sum of the conversions by gasification reactions. When applied to the simulation data from the non-reactive boundary-layer case (see Fig. 9), the mutually exclusive conversion was found to be determined by oxygen. In Figs. 11 and 12, the predictions using the mutually exclusive approach do not agree with the simulated results from the proposed model. The conversion rate is practically constant, regardless of char gasification reactivity. The mutually exclusive rate curve would have a different shape if it were determined by the sum of conversions by gasification reactions. In practice, there is likely to be proportionally more carbon dioxide and water vapor and less oxygen in the bulk gas. For purposes of discussion, this kind of situation is illustrated in Fig. 13. The conversions in Fig. 13 were calculated from the original conversions shown in Fig. 9 by halving the conversion by oxygen and doubling the conversions by gasification reactions.

In Fig. 13, the mutually exclusive conversion curve can be seen to have a shape more closely resembling those predicted by the proposed model (Figs. 11 and 12). However, the mutually exclusive concept does not predict a decrease in conversion at zero gasification because it does not take into account that oxygen is consumed by the carbon monoxide produced from direct oxidation. Another conflicting feature of Fig. 13 is that for mass transfer-limited burning, the mutually exclusive conversion curve is below that representing the
sum of conversion by all three oxidants. According to the proposed model, there should be no decrease if boundary-layer chemistry only is considered (Fig. 12). It could also be argued that when considering the impact of increased boundary-layer temperature on net mass transfer, there should be a decrease in overall conversion under mass transfer-limited conditions. However, at least in the simulations shown in Fig. 11, the magnitude of the conversion decrease predicted by the mutually exclusive concept does not agree with the results obtained here. To achieve an updated simplified model, boundary-layer theory in combination with an energy balance could possibly provide a more rigorous approach.

**Predicted vs. experimental results**

Simulations of IPST Cases 1–4 were carried out under the assumptions of a non-reactive and a reactive boundary layer. Figure 14 presents the model predictions. The predicted conversion rates increased from Case 1 to Case 4 because of the increase in total oxygen available in the gas jet. In Cases 1–4, the gas-phase reactions decreased the impact on predicted carbon conversion. In Fig. 14, the predicted bed temperatures decreased from Case 1 to Case 4 because carbon conversion took place increasingly through endothermic reactions. In Cases 1–4, gas-phase reactions had an increasing impact on predicted bed temperatures.

As for carbon oxides, carbon monoxide was the major char-carbon conversion product in predictions with a non-reactive boundary layer. Even part of the carbon dioxide from sulphate reduction reacted further with carbon monoxide through the gasification reaction. The smaller percentages of carbon monoxide in Cases 2 and 4, as compared to Cases 1 and 3, were due to the presence of carbon dioxide in the gas jet. When the boundary layer was reactive, all carbon monoxide was oxidized to carbon dioxide in Cases 1–4.

When comparing the predicted and experimental data in Fig. 14, it can be observed that the predicted conversion rates are generally lower than the experimental ones. One possible explanation is the model assumption of a flat and smooth bed. Most likely the experimental bed was non-flat and rough. Both these factors can be expected to contribute to higher burning rates. Figure 14 also shows that the predicted temperatures were on average of similar magnitude to the experimental values, but that the predicted carbon oxides ratio agreed with experimental data only in Cases 3 and 4, where the boundary layer was assumed to be reactive. In Cases 3 and 4, the predictions were in reasonable agreement with experimental data, apart from the absolute magnitude of the conversions. The model correctly predicted the trend of increasing conversion and decreasing temperature from Case 3 to Case 4.

The agreement between predicted and experimental carbon oxide values in Cases 3 and 4 indicates that the model predicted the boundary-layer chemistry correctly in these cases. By contrast, there is a lack of agreement between Cases 1 and 2. The predicted percentages of carbon monoxide were too high for an assumed non-reactive boundary layer and too low for an assumed reactive boundary layer. One motivation for testing the chemical mechanism also in Cases 1 and 2 was the hypothesis that hydrogen released from char conversion would be sufficient to initiate gas-phase reactions. The discrepancy indicates that this was an unrealistic assumption and that the chemical mechanism used is unsuitable for cases where there is no water vapor in the bulk gas.

The experimental carbon oxides data suggest that in Cases 1 and 2, some oxidation might have occurred inside the boundary layer, although to a lesser extent than when water vapor was present. \( \text{CO} + \text{O}_2 \rightarrow \text{CO}_2 + \text{O} \) is an elementary reaction relevant to the oxidation of carbon monoxide in the absence of water vapor. Simulations of Cases 1 and 2 using this reaction were carried out. These indicated that practically no carbon monoxide was oxidized in the boundary layer. The results were very similar to those for a non-reactive boundary layer. More work is needed if better agreement with experimental data is desired. However, a better understanding of the relevant gas-phase chemistry in the absence of water vapor is perhaps not of critical importance when considering potential full-scale simulations. In a recovery furnace, there will always be some water vapor present in the bulk gas.

**SUMMARY AND CONCLUSIONS**

A mathematical model of char-bed burning in the presence of a chemically reactive boundary layer has been presented. Simulations of the IPST char-burning experiments were carried out. The following conclusions can be drawn:

- Boundary-layer reactions consume oxygen, which has a net exothermic impact on overall combustion, increasing both gas-phase and char-bed temperatures;
- oxygen in the bulk gas contributes to carbon conversion, although carbon conversion involving gas-phase reactants can occur entirely through gasification reactions;
- overall carbon conversion can be decreased either because of the direct effect of boundary-layer gas-phase chemistry or because combustion inside the boundary layer affects mass transfer through this layer;
- the mutually exclusive concept of comparing conversion rates by direct oxidation and gasification needs to be revised if simplified calculations of overall burning rate during combined oxidation and gasification are to be consistent with the numerical simulations reported in this paper;
- when compared to experimental data, the model predictions are in reasonable quantitative agreement on the effect of gas-phase reactions on burning rate, bed temperature, and carbon oxides formation during simultaneous oxidation and gasification;
- the current model is unsuitable for simulations of bed burning in the absence of water vapor.
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REFERENCES


NOTATION

\begin{tabular}{|l|l|}
\hline
\(A_s\) & specific surface area of char \\
\(G\) & concentration \\
\(D_v\) & binary diffusion coefficient \\
\(D_{gas}\) & diffusivity of species \(i\) in gas mixture \\
\(D_{eff}\) & effective diffusivity of species \(i\) in char bed \\
\(\varepsilon\) & char porosity \\
\(\phi\) & Thiele modulus \\
\(\bar{h} = \eta \tau L\) & kinetic rate coefficient \\
\(L\) & char-bed burning active-layer thickness \\
\(\lambda\) & thermal conductivity of gas \\
\(m_{tu}\) & mass transfer coefficient \\
\(t_{tu}\) & turbulent viscosity \\
\(\phi_{m}\) & molecular viscosity \\
\(N\) & chemical species flux \\
\(Na_{Na}\) & sodium content of char \\
\hline
\(\eta\) & effectiveness factor \\
\(R\) & reaction rate \\
\(R_{act}\) & reaction rate given by chemical kinetics \\
\(R_{tur}\) & reaction rate given by turbulent dissipation \\
\(R_{tur,m}\) & modified turbulent dissipation reaction rate \\
\(R_{CO_2}\) & rate of carbon gasification by \(CO_2\) \\
\(R_{H_2O}\) & rate of carbon gasification by \(H_2O\) \\
\(R_{CO}\) & rate of carbon gasification by \(CO_2\) or \(H_2O\) \\
\(R_{C-x}/A\) & carbon gasification per unit char-bed surface area \\
\(R_{S-x}/A\) & rate of sulphate reduction \\
\(R_{SO_2}/A\) & sulphate reduction per unit char-bed surface area \\
\(r\) & intrinsic reaction rate \\
\(\rho_{char}\) & char-bed bulk density \\
\(\tau\) & tortuosity factor \\
\(x\) & distance \\
\(y\) & mole fraction \\
\hline
\end{tabular}
Pulp and paper companies need new methodologies to evaluate capital spending opportunities using improved cost models and available cost and process data. This paper presents an application of the operations-driven cost model presented in Part I (Laflamme-Mayer, M., Janssen, M., Stuart, P., Journal of Science & Technology for Forest Products and Processes, 1:1 (2011)) to a retrofit design case study defined at an existing integrated newsprint mill. The paper outlines the structure of the cost model and discusses its application to the analysis of a set of design alternatives. The operations-driven cost modelling approach significantly increases the granularity and transparency of the results obtained from a techno-economic study and permits the examination of critical design variables and operating variants.

MATTY JANSSEN, PAUL NALIWAJKA, PAUL R. STUART*

INTRODUCTION

This paper is the second in a series that outlines the theoretical foundations and application of a new operations-driven cost modelling approach [1,2]. More specifically, it seeks to demonstrate the application of the operations-driven cost modelling approach to a retrofit design problem using the data and structure of an ABC-like cost accounting system. The objectives of this study were:

- To demonstrate the value of the operations-driven cost modelling approach in the evaluation of design alternatives for design decision making.
- To use the cost model to obtain insight into design alternatives by considering operating variants (established by, e.g., changing the production capacity of a design alternative or varying the steam production or demand of the mill) based on energy efficiency studies and analysis of production capacity change.
- To identify the benefits of using this approach compared to a rigorous but “classical” techno-economic analysis.

The case study considers the implementation of increased DIP pulp production and cogeneration at an existing integrated newsprint mill [3].

Retrofit Process Design and Cost modelling in the Pulp and Paper Industry

One of the opportunities stemming from the effective use of data from Information Management Systems (IMS) at pulp and paper mills lies within the field of retrofit process design. By using more detailed better-structured models, analytical tools, and methodologies with the data gathered by an IMS, better decisions can be design alternative for investment [4]. Manufacturing costs for retrofit design alternatives can be compared to current manufacturing costs using mill process and cost data in financial spreadsheet models, and the resulting NPVs for the alternatives can be calculated. Furthermore, more advanced analysis such as Monte Carlo risk analysis can be carried out to assess risks associated with key project variables and parameters [3,5].

Classical techno-economic studies in the pulp and paper industry typically use volume-based cost accounting data as
the basis for modelling costs. These studies yield good information for decision-making for many design objectives, but can, however, lack the granularity (detail) and transparency (understanding of cause and effect) that would permit a more careful evaluation of design alternatives when required. Volume-based cost accounting was developed primarily for discrete manufacturing processes. The resource consumptions and activities needed to manufacture a product with such a process are known from routing information and bills of materials and can therefore be traced directly to the product, as is done in volume-based cost accounting. Continuous processes, such as pulp and paper processes, can be divided into main-line processing operations and converting and finishing operations. In main-line processing operations, costs are generated by the production process and its operating conditions. For instance, activities such as maintenance (which is a typical overhead cost) are linked to the production process, and the operating conditions determine the resource consumptions. This implies that cost-driver relationships for processing operations should be based on process-related aspects. To increase understanding of the cost implications of a particular retrofit design alternative to a continuous process, it is therefore of interest to use a cost accounting method that can capture these process-related aspects. Activity-based costing (ABC) is such a method and will increase both the transparency and the granularity of a cost model compared to a volume-based cost model.

Using ABC principles, cost accounting data can be systematically used for such applications as tracking production costs and cost variances to reduce production and quality variability [6] and the evaluation of retrofit design alternatives by systematically reconciling process and cost data. When a mill has an ABC-like system in place, opportunities for more advanced and sophisticated analysis can be exploited, such as decision-making for capital spending through sensitivity analysis, incremental and marginal cost analysis, or risk analysis. A cost model based on ABC can thus extract more relevant information for operational or design decision-making from the cost and process data that are available at a mill. The first paper in this series included a more in-depth discussion of ABC [1].

Integrated newsprint mill energy considerations and costs

Although a number of techniques are available for reducing TMP (thermo-mechanical pulping) energy consumption, they generally yield only marginal reductions compared to the implementation of de-inked pulp (DIP) production to replace TMP (assuming a constant production rate). However, by decreasing TMP production, the production of steam from the TMP plant (used principally for paper drying operations) is reduced. The mill must compensate for this loss of steam by increasing steam production in the boilers. Consequently, this increase can give rise to further capital spending requirements, leading to consideration of cogeneration at the mill. Cogeneration is the combined production of electrical (or mechanical) and thermal energy from the same primary energy source [7]. Energy efficiency studies can be carried out to optimize the profitability of a design alternative by, for example, reducing process steam demand. Pinch analysis can be used for thermal optimization and for definition and organization of energy efficiency projects by reducing both energy and water use at a mill [8,9,10]. Furthermore, marginal energy cost analysis helps to identify the operating conditions at which maximum profitability can be achieved.

In marginal economics, both incremental manufacturing costs and revenues are seen as variable. This results in a more realistic view of how costs per unit produced change and may lead to the observation that unit manufacturing costs decrease at first, but then at some level start to rise as production increases (Fig. 1). There is therefore an optimum for capacity utilization [11]. Marginal cost analysis identifies operating scenarios that maximize the cash flow for a given investment and for different operating conditions. Furthermore, the use of production functions to characterize the (non-linear) resource behaviour of operations provides a more accurate view of manufacturing costs [12].

EXISTING MILL AND DESIGN ALTERNATIVES

Existing mill design

The existing mill on which this study is based consists of the following production units:

- 4 newsprint machines with a total average production of 1100 tonnes/day of newsprint;
- 2 TMP lines with a total average production of 925 tonnes/day of pulp;
- A DIP plant with a total average production of 175 tonnes/day of pulp, where 85% of the wastepaper used is old newspaper (ONP) and 15% is old magazine paper (OMG).

Furthermore, the following supporting processes are part of the base-case mill design:

- A wastewater treatment plant processing 50,000 m³/day;
- A boiler plant producing 7850 GJ/day of steam;
- A back-pressure turbine generating only 0.4% of the total mill electricity demand.
De-inking and cogeneration design alternatives

The DIP plant process designs considered in the study would increase DIP production to either 550 tonnes/day (50%) or 1100 tonnes/day (100%) (Table 1). Both one-loop and two-loop DIP technologies were considered. A one-loop DIP system is a system that processes the recycled paper in one alkaline cleaning stage. A two-loop system has an additional second cleaning loop that operates under acidic conditions, making it a more rigorous cleaning process. The one-loop de-inked plant technology is the typical technology used in North America, and its capital cost is lower than that of a two-loop system with the same capacity. However, two-loop technology includes additional equipment that can compensate for the expected quality loss of recycled paper in the future [13]. The cogeneration designs studied here have the following characteristics (Table 2 and Fig. 2):

- Increase of biomass combustion capacity;
- Reactivation of idled turbines and implementation of new back-pressure turbines or condensing turbines.

In total, 18 alternatives were analyzed in this case study by considering all combinations of the 6 DIP and 3 cogeneration designs. The following naming convention for the design alternatives was used: Alternative {DIP design 1 to 6 as per Table 1}-{Cogeneration design A, B, or C as per Table 2}.

METHODOLOGY

After developing the operations-driven cost model for existing mill processes, the study methodology consisted of the following steps for each of the design alternatives (Fig. 3):

1. Calculation of mass and energy balances for each process flowsheet;
2. Calculation of total capital costs;
3. Modelling and calculation of operating costs for the design alternatives and operating variants;
4. Profitability calculations for the evaluation of the design alternatives and operating variants.

These four steps are discussed in more detail below. The model was developed using the Impact: EDC™ software package from 3C Software Inc. [14].

Capital cost estimates and mass and energy balances

The total capital cost and the mass and energy balance models were constructed as described in Janssen et al. [3], and the same hypothetical mill and design hypotheses.
were assumed (steps 1 and 2 in Fig. 3). The mass and energy balance model was used to examine the impact of variations in steam generation in the boilers. Within this model, the amount of cogenerated power was also calculated. The results of these models for all 18 design alternatives were then connected to the operations-driven cost model.

Operations-driven cost model

Process design data generated by the mass and energy balances were used as inputs to the operations-driven cost model [1] (step 3 in Fig. 3). For calculation of variable costs, Process Work Centres (PWCs) were used to represent different mill processes. The PWCs were divided into production and support PWCs (Fig. 4). The overhead costs were calculated in the Overheads Work Centre (OWC). The data used in this study (which were taken from the accounting system at an operating integrated newsprint

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**TABLE 1  DIP plant designs**

<table>
<thead>
<tr>
<th>Design</th>
<th>Technology and capacity</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>New 550 t/d DIP plant, 1-loop</td>
</tr>
<tr>
<td>2</td>
<td>New 550 t/d DIP plant, 2-loop</td>
</tr>
<tr>
<td>3</td>
<td>New 1100 t/d DIP plant, 1-loop</td>
</tr>
<tr>
<td>4</td>
<td>New 1100 t/d DIP plant, 2-loop</td>
</tr>
<tr>
<td>5</td>
<td>Increase to 550 t/d by adding a second line to the existing plant, 1-loop</td>
</tr>
<tr>
<td>6</td>
<td>Increase to 550 t/d by adding a second line to the existing plant, 2-loop</td>
</tr>
</tbody>
</table>

**TABLE 2  Cogeneration designs**

<table>
<thead>
<tr>
<th>Design</th>
<th>Description</th>
<th>Steam production</th>
</tr>
</thead>
<tbody>
<tr>
<td>Existing mill</td>
<td>All boilers operate at high pressure (HP) and only one turbine is in service</td>
<td>At HP: 90,000 tonnes/year</td>
</tr>
<tr>
<td>A</td>
<td>One natural gas boiler is converted to burn wood waste, and existing backpressure turbines kept in service.</td>
<td>At HP: 150,000 tonnes/year</td>
</tr>
<tr>
<td>B</td>
<td>New wood waste boiler (at very high pressure (VHP)) is installed. Three out of six boilers are upgraded to VHP operation. New backpressure turbine added.</td>
<td>At VHP: 130,000 tonnes/year, At HP: 25,000 tonnes/year</td>
</tr>
<tr>
<td>C</td>
<td>New condensing turbine installed.</td>
<td>At HP: 100,000 tonnes/year</td>
</tr>
</tbody>
</table>

---

**Fig. 3 - Operations-driven cost modelling approach.**
The traditional area contributions provided detailed information about the indirect manufacturing cost per PWC and could therefore be directly related to the PWCs. Specifically allocated non-manufacturing costs were distributed equally over the PWCs in a second-level allocation. Some overhead costs varied for different design alternatives and operating variants and were modeled in the OWC. For instance, due to changes in production volumes, the labour costs and headcount in the TMP and DIP PWCs were different. In the case study, the following PWCs had varying overhead costs: TMP, chip handling, DIP, turbines, and boilers.

**Process design evaluation (step 4 in Fig. 3)**

Evaluation of design alternatives: Various profitability metrics were calculated to evaluate the feasibility of the design alternatives, including net present value (NPV), internal rate of return (IRR), return on investment (ROI) and cumulated economic value added (EVA). The financial parameters were the same as used in Janssen et al. [3]. EVA is an estimate of the wealth creation potential from capital investments. It attempts to capture the true economic profitability of a company or project and accounts for the cost of capital [15]. Calculation of NPV is based only on cash flow, whereas EVA also takes into account the riskiness of an investment. Sensitivity analyses were carried out for electricity and natural gas prices.

Marginal energy cost analysis: In this analysis, the marginal steam cost (in $/GJ) and the incremental and marginal cost of power produced by cogeneration (in $/MWh) were calculated. This was done by first varying the steam production in the steam plant over a range of values. Each of these values represents an operating variant. Next, the total steam cost and turbine steam cost (i.e., the cost of steam used by turbines for cogeneration) were calculated using the steam price:

\[
\text{Total steam cost} = \text{Steam produced} \times \text{Steam price}
\]

\[
\text{Turbine steam cost} = \text{Turbine steam use} \times \text{Steam price}
\]

The steam price (The term “steam price” is used to indicate that the generated steam is sold internally by the Boilers PWC to the other PWCs. The cost of fuels used to generate the steam is thus routed to the PWCs where the steam is used.) was calculated based on the fuels used. Using difference equations, the marginal steam cost and marginal cost of generated power can be calculated as follows:

\[
\text{Marginal steam cost} = \frac{\text{Total steam cost}_{i+1} - \text{Total steam cost}_i}{\text{Steam produced}_{i+1} - \text{Steam produced}_i}
\]

\[
\text{Marginal cost of generated power} = \frac{\text{Turbine steam cost}_{i+1} - \text{Turbine steam cost}_i}{\text{Power generated}_{i+1} - \text{Power generated}_i}
\]

where \(i\) refers to an operating variant.

The incremental cost of generated power for each alternative was calculated relative to the existing mill:

\[
\text{Incremental cost of generated power} = \frac{\text{Turbine steam cost}_{\text{alt.}} - \text{Turbine steam cost}_{\text{ref.}}}{\text{Power generated}_{\text{alt.}} - \text{Power generated}_{\text{ref.}}}
\]

where alt. refers to the retrofit design alternative under consideration.

Energy efficiency: The impact of increased energy efficiency in the process was considered in certain scenarios. This was done by reducing the steam use of the mill over a range of values up to \(6.45 \times 10^5\) GJ/y, corresponding to a maximum reduction of 20% for the 100% DIP alternatives and of 25% for the 50% DIP...
alternatives, followed by calculation of the marginal steam cost, cost of generated power, and profitability.

Production capacity change: Production functions were established to add non-linear resource behaviour resulting from a change in paper production. Functions for electricity and steam use on the paper machines were determined using the following assumptions (Fig.s 5 and 6):

- Base load steam: 30% of nominal use;
- Base load electricity: 70% of nominal use;
- Steam use for drying decreases by 1% per 10 tons of production increase.

The base load refers to the amount of steam or electricity used by the paper machine when no paper is produced. Causes of non-linearity in the paper machines include the steam condenser efficiency and the efficiencies of paper machine drives and pumps.

Production functions for the efficiency of the cogeneration designs were then determined (Fig. 7). Because no data were available for the boilers separately, these functions are composite functions describing all the boilers in a design. A quadratic relationship between steam production and efficiency was assumed [16], with maximum efficiency set to 75% for base-case steam production. This non-linear behaviour may be caused by moisture levels in the fuel and air, incomplete combustion, combustion of hydrogen (to water), and radiation [17]. Linear relationships were assumed with pulp production for electricity use in the TMP plant and with steam production for power generation in the turbines [17]. Furthermore, it was assumed that electricity use in the 100% DIP plants remains constant with changing de-inked pulp production because a constant volumetric rate is maintained.

RESULTS AND DISCUSSION

Manufacturing costs and profitability

The profitability of the design alternatives was calculated under the following...
conditions:

- Declining balance method for depreciation with a fixed depreciation rate [18],
- Investment tax credit of $10/MWh of electricity generated based on renewable fuels [19],
- Sale of generated electricity to the grid at the nominal electricity price plus a 50% premium. This premium stimulates mills to sell their cogenerated power.

The manufacturing costs were split into direct and overhead costs and specified per PWC (Fig. 8). A negative value for a PWC cost indicates that this PWC is a profit centre. For instance, for alternative 1-A, the manufacturing cost is $(310 - 37) + (217 - 18) = 472 \$/tonne.$ The turbine work centre is an important profit centre because of sales of generated electricity to the grid. Therefore, the turbine PWC direct cost for each alternative is negative. The cost of fibre is the most significant cost for the TMP and DIP PWCs and depends on the implemented DIP capacity (50% or 100%) as well as the yield difference between the one-loop and two-loop designs (85% vs. 92%). Furthermore, there is a difference in steam consumption: the one-loop design uses 1.25 GJ/t and the two-loop design 2.25 GJ/t of waste paper used. This is caused by the use of extra steam in deflocculation for better pulp cleaning in the two-loop design. The deflocculation unit uses steam and electrical energy to break down dirt and stickies remaining in the pulp for easier removal in the second loop. The variation in direct cost for the paper mill PWC can be explained primarily by the variation in steam price because the paper mill is a large steam consumer. The steam price depends on the wood waste capacity of each cogeneration design and the amount of natural gas and sludge used for steam generation. One advantage of the operations-driven cost modelling approach is that a higher manufacturing cost for the paper mill PWC can be clearly traced back to higher steam prices resulting from an increase in the cost of natural gas. Using a volume-based cost model, this increase would only be reflected in the final product.
manufacturing cost. The model would not be capable of determining directly which process unit is responsible for this increase and why the increase has occurred.

The overhead costs for each alternative show only small variations compared to direct costs (Fig. 8). These variations occur because of:

- Differences in the investment tax credit (ITC); the more power is generated based on renewable fuels, the more ITC is received;
- Differences in maintenance material and labour, operating labour, and supply costs and changes in headcount;
- Differences in the cost of mill heating, which varies with steam price.

For each PWC, the “assigned overhead charge” is the non-manufacturing cost that was assigned directly to the cost object, the newsprint paper produced. This cost is negative because it contains costs that are transferred to other facilities at the mill site. Using the operations-driven costing approach, the PWCs that contribute to a manufacturing cost change can be identified. At a more detailed level, the change in cost for each activity in these PWCs can be assessed.

Based on the results of the profitability analysis (Fig. 9), the alternatives that had a positive value for all the criteria were retained and further analyzed. Alternative 3-A was most profitable, with an NPV of 82.3 M$ and an IRR of 8.1%.

Sensitivity analyses were carried out to assess the impact of electricity and natural gas prices on the profitability of the design alternatives (results not shown). The natural gas and electricity consumption levels are lower for the design alternatives than for the existing mill, and therefore increased energy prices have a positive impact on profitability. Both electricity and natural gas prices influence the steam price and the incremental cost of generated power (Eq. 5) (Fig. 10). An electricity price change has a minor effect on both, even in the extreme case of a 100% change. However, an increase in the natural gas price results in a more significant change. The incremental cost of generated power can be reduced, change only a little, or increase depending on the steam price increase for the existing mill and the alternative.

**Marginal energy cost analysis**

Figure 11a shows the impact of changes in fuel on steam cost by plotting marginal and average steam costs at different steam production rates. For alternatives with cogeneration designs A and B, the marginal steam cost is lower than the average steam
cost when steam production is from wood waste (lower steam production rates). After natural gas is added to the fuel mix, the marginal steam cost is significantly higher than the average steam cost. For instance, the average steam cost in alternative 3-A is $2.15/GJ until natural gas is added, after which the marginal cost jumps to $10.50/GJ. Only alternative 3-C does not display a jump in the marginal steam cost and has a marginal steam cost higher than the average steam cost at all steam production rates because of natural gas use at these rates. The marginal cost of generated power shows a similar trend (Fig. 11b).

For the alternatives with cogeneration designs A and B, an optimal NPV is identified as a function of variant parameters, specifically when wood waste use is maximized and natural gas use is minimized. These results reflect the outcomes of the marginal energy cost analyses, i.e., these alternatives would start to lose money as soon as natural gas is required (Fig. 11c).

Energy efficiency
Increasing process energy efficiency does not result in a significant impact on marginal steam cost, nor on the incremental and marginal costs of generated power. However, the impact on profitability is more profound (Fig. 12). If steam production in the boilers is decreased by the same quantity as is conserved by the higher energy efficiency, then the NPV decreases significantly. The decrease in steam production leads to lower electricity generation, and therefore the mill receives less revenue from sale of this power. For instance, the NPV of alternative 3-A decreases from $82.3 million to $71.4 million when steam production in the boiler plant is decreased by 15%. The operations-driven approach can readily quantify such changes in profitability. If steam production
in the boilers remains at the base operating level, the NPV increases with decreasing process steam use. For greatest profitability, the cogeneration potential should remain as high as possible while maintaining an optimal fuel mix.

**Production capacity change**

The paper-making process exhibits non-linear behaviour with regard to electricity and steam use by the paper machines and to boiler efficiency. The manufacturing costs for various production capacities have been calculated using the production functions (shown earlier in Figs. 5, 6, and 7) and using constant values for those production variables. The calculations were carried out for design alternatives 3-A and 5-A. Using production functions leads to different manufacturing costs (Fig. 13). This difference is amplified as the production rate deviates from the base design specification (in this case, 1100 FMT/day for all paper machines). The overhead costs were not varied in these calculations.

The marginal manufacturing cost varied over the production range, resulting in non-linear behaviour of the manufacturing cost (Fig. 14) with production rate. This marginal cost was calculated using the following equations:

\[
\text{Marginal manufacturing cost} = \frac{\text{Total manufacturing cost} - \text{Total manufacturing cost}}{\text{Production rate} - \text{Production rate}}
\]

where \(i\) refers to an operating variant.

The marginal manufacturing cost over the production range is higher for alternative 5-A than for the other scenarios. This implies that there is a significant impact of TMP operating costs (i.e., electricity use) on the marginal cost of the alternatives, because alternative 3-A produces no TMP pulp. For both alternatives, the marginal manufacturing costs stay well below the average manufacturing cost (Fig. 13), which indicates that increased paper production...

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**Fig. 13 - Manufacturing cost for: a. alternative 3-A; and b. alternative 5-A.**
production leads to increased earnings at all production rates.

**Comparison with conventional techno-economic approach**

Janssen *et al.* [3] presented a techno-economic study for the same design problem as presented in this study. By comparing the operations-driven cost modelling approach used in this study with that using classical accounting data in the earlier techno-economic study, it is evident that:

- The transparency and granularity of the results increases because of the use of the ABC-like approach. Resources are related to cost objects through the activities performed in the defined *Process Work Centres*. In techno-economic studies, the resources are directly related to the cost object, i.e., the resource costs are paid for by the whole mill instead of by the process unit where the resource is used.
- The operations-driven approach enables detailed integration and reconciliation of the process and cost data in one model. The *PWCs* can be used to study a process at any level of detail, depending on the objective of the study and the availability and accuracy of data. This permits a more efficient and accurate description of changes in direct and overhead costs based on design changes. In a techno-economic study, the data are not captured in such a framework, and it is therefore less clear how to link design changes to cost changes.
- The current approach integrates the calculation of design variables that change with resource price, e.g., steam price. The costs of resource use can be traced back to the *PWCs* that use these resources indirectly, e.g., changing steam cost for the paper mill *PWC* as a function of varying natural gas price. Such changes are not so obviously taken into account in a techno-economic study.

**CONCLUSIONS AND IMPLICATIONS**

This study has sought to apply an operations-driven cost modelling approach to a large-scale retrofit design problem. It has.

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**Fig. 14 - Marginal manufacturing cost for:**

- **a. alternative 3-A;** and
- **b. alternative 5-A.**
considered the implementation of increased de-inked pulp production and cogeneration at an integrated newsprint mill.

First, the mass and energy balances and capital costs were calculated for all design alternatives. Next, the operations-driven cost model was used to calculate the operating costs and profitability of these alternatives. The profitable alternatives were identified, and marginal cost analyses and energy efficiency studies were carried out to analyze these alternatives in more depth.

The profitability analysis showed that a 100% DIP one-loop alternative was the most profitable (alternative 3-A) and that none of the two-loop DIP design alternatives was profitable. Marginal cost analysis quantified the negative effect of natural gas use on profitability due to its high price. Furthermore, an energy efficiency study showed that profitability increases only when maintaining steam production in the boiler plant while increasing the energy efficiency of the mill processes. This information can be used in the design decision-making process to account for operating conditions that are different from the design specifications.

The proposed cost modelling approach is better able than other methods to quantify cost implications of retrofit design changes due to its focus on mill processes. The approach readily permits carrying out marginal cost analysis and energy efficiency studies because of its ability to calculate cost variations by evaluating operating variables such as steam production rate.

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REFERENCES

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