Upgrading of waste streams into fuel by means of drying with low temperature excess heat

a techno-economic evaluation

Tiina Myllymaa

Upgrading of waste streams into fuel by means of drying with low temperature excess heat

a techno-economic evaluation

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A doctoral dissertation completed for the degree of Doctor of Science (Technology) to be defended, with the permission of the Aalto University School of Engineering, at a public examination held at the lecture hall K216 of the school on 13 December 2019 at 12 noon.

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Aalto University publication series

DOCTORAL DISSERTATIONS 215/2019

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ISBN 978-952-60-8833-4 (printed)

ISBN 978-952-60-8834-1 (pdf)

ISSN 1799-4934 (printed)

ISSN 1799-4942 (pdf)

<http://urn.fi/URN:ISBN:978-952-60-8834-1>

Images: Akvarel painting of front cover: Terhi Järvinen

Unigrafia Oy

Helsinki 2019

Finland

Printed matter
4041-0619

Author

Tiina Myllymaa

Name of the doctoral dissertation

Upgrading of waste streams into fuel by means of drying with low temperature excess heat - a techno-economic evaluation

Publisher School of Engineering

Unit Mechanical Engineering

Series Aalto University publication series DOCTORAL DISSERTATIONS 215/2019

Field of research Energy Technology

Manuscript submitted 10 May 2019

Date of the defence 13 December 2019

Permission for public defence granted (date) 28 October 2019

Language English

☐ **Monograph**

☒ **Article dissertation**

☐ **Essay dissertation**

Abstract

The use of industrial excess heat in drying of combustible waste streams enhances the energy efficiency of combustion process and reduces emissions. A suitable dryer for wastes is a belt dryer when low air temperatures are used. However, the total drying costs may vary considerably depending on how different design parameters are chosen. This dissertation studies the effects of the design parameters of the belt dryer on total drying costs when different waste streams are dried using low temperature excess heat. Temperature is usually lower than 100-120 °C. The waste materials included are woodchips, bark, and mixture of soot sludge and sawdust. The influence is assessed by changing the following parameters: bed height, drying air temperature, air velocity, and initial/final moisture contents of the material. In order to produce a techno-economic evaluation, a series of experimental tests have been conducted to gain information on drying kinetics of the streams in fixed beds. Based on the experimental data, an Excel-based programme was developed to determine the costs.

Drying kinetics studies show that the air temperature has a great impact on the drying time for all materials; the lower the temperature, the longer the drying time needed. For example, for woodchips, the drying time is ~two times longer when the temperature decreases from 70 to 50 °C (300 mm). Results indicate that there is an optimum value for air velocity, after which increase of the velocity will no longer remarkably decrease the drying time. Results also indicate that the most optimal volume ratio for the sludge and sawdust is 50%:50% and the most optimal bed height is 200 mm.

The economic calculations show that the highest bed height and the highest air temperature (bark: 250 mm, 110 °C; woodchips: 500 mm, 90 °C) yield the lowest drying costs, regardless of the change of moisture content, when the price of the heat is the same for every air temperature. For the sludge and sawdust mix, the lowest costs are gained with the highest air temperature (100 °C). The optimal bed height of the mix depends on the material and its initial moisture content. The higher the moisture content, the lower the optimal bed height is.

When materials are dried, the availability of heat sources at different temperatures may be limited. In that case, it may be reasonable to use different air temperatures in successive drying blocks. On the basis of block studies, it seems to be more economic to use lower air temperatures in the first blocks when material is moist, and higher air temperatures in the last blocks when the moisture content of material is lower if the heat price is the same for both temperature levels. Air velocity should be the same over the entire dryer. In general, the results of this dissertation shows that it is important to pay attention to the main design parameters to optimize total drying costs.

Keywords drying costs, belt dryer, waste heat, woody biomass, bark, woodchips, sawdust, soot sludge

ISBN (printed) 978-952-60-8833-4

ISBN (pdf) 978-952-60-8834-1

ISSN (printed) 1799-4934

ISSN (pdf) 1799-4942

Location of publisher Helsinki

Location of printing Helsinki **Year** 2019

Pages 148

urn <http://urn.fi/URN:ISBN:978-952-60-8834-1>

Tekijä

Tiina Myllymaa

Väitöskirjan nimi

Jätevirtojen jalostaminen polttoaineeksi kuivauksella käyttäen matalalämpötilaista ylijäämälämpöä - tekno-taloudellinen arviointi

Julkaisija Insinööritieteiden korkeakoulu**Yksikkö** Konetekniikan laitos**Sarja** Aalto University publication series DOCTORAL DISSERTATIONS 215/2019**Tutkimusala** Energiatekniikka**Käsikirjoituksen pvm** 10.05.2019**Väitöspäivä** 13.12.2019**Väittelyluvan myöntämispäivä** 28.10.2019**Kieli** Englanti☐ **Monografia**☒ **Artikkeliväitöskirja**☐ **Esseeväitöskirja****Tiivistelmä**

Teollisuuden ylijäämälämmön käyttö palavien jätevirtojen kuivauksessa lisää palamisprosessin energiatehokkuutta ja vähentää päästöjä. Sopiva kuivuri jätteille on hihnakuivuri, kun käytetään alhaisia ilman lämpötiloja. Kuivaukset voivat kuitenkin vaihdella huomattavasti riippuen siitä, miten eri suunnitteluparametrit valitaan. Tässä väitöskirjassa tutkitaan hihnakuivurin suunnitteluparametrien vaikutuksia kokonaiskuivaukskustannuksiin, kun eri jätevirrat kuivataan matalalämpötilaisella ylijäämälämmöllä (alle 100-120 °C). Työhön sisällytetyt materiaalit ovat puuhake, kuori ja nokilietteen ja sahanpurun seos. Vaikutusta arvioidaan muuttamalla seuraavia parametreja: petikorkeus, kuivausilman lämpötila, ilman nopeus ja materiaalin alku-/loppukosteuspitoisuus. Tekno-taloudellisen arvioinnin aikaansaamiseksi on tehty useita kokeellisia testejä, jotta ollaan saatu tietoa jätevirtojen kuivauksinetiikasta kiinteässä pedissä. Koetulosten perusteella kehitettiin Excel-ohjelma kustannusten määrittämiseksi.

Kuivauksinetiikkatutkimukset osoittavat, että ilman lämpötilalla on suuri vaikutus materiaalien kuivausaikaan; mitä alhaisempi lämpötila, sitä pidempi kuivausaika. Esimerkiksi hakkeella kuivausaika on noin kaksi kertaa pidempi, kun lämpötila laskee 70:sta 50 °C:een (300 mm). Tulokset osoittavat, että ilman nopeudelle on optimi, minkä jälkeen nopeuden lisäys ei enää vähennä huomattavasti kuivausaikaa. Tulokset osoittavat myös, että lietteen ja purun optimaalinen tilavuussuhde on 50%:50% ja pedin korkeus on 200 mm.

Taloudelliset laskelmat osoittavat, että korkein petikorkeus ja korkein ilman lämpötila (kuori: 250 mm, 110 °C; hake: 500 mm, 90 °C) antavat alhaisimmat kuivaukskustannukset, riippumatta kosteuspitoisuuden muutoksesta, kun lämmön hinta on sama kaikille ilman lämpötiloille. Lietteen ja purun seokselle alhaisimmat kustannukset saadaan korkeimmalla ilman lämpötilalla (100 °C). Seoksen optimi petikorkeus riippuu materiaalista ja sen alkukosteudesta. Mitä suurempi kosteus pitoisuus on, sitä pienempi on optimaalinen pedin korkeus.

Lämmönlähteiden saatavuus eri lämpötiloissa voi olla rajoitettu kuivauksessa. Tällöin voi olla järkevää käyttää eri ilman lämpötiloja peräkkäisissä kuivurilohkoissa. Tutkimuksen perusteella on taloudellisempaa käyttää alhaisempaa ilman lämpötilaa ensimmäisessä lohossa, kun materiaali on kostea, ja korkeampaa ilman lämpötilaa seuraavassa lohossa, kun materiaalin kosteus on pienempi, jos lämmön hinta on sama molemmille lämpötiloille. Ilman nopeuden tulisi olla sama koko kuivurissa.

Yleisesti tulokset osoittavat, että on tärkeää kiinnittää huomiota pääsuunnitteluparametreihin kokonaiskuivausmenojen optimoimiseksi.

Avainsanat kuivaukskustannukset, hihnakuivuri, jätelämpö, puubiomassa, kuori, puuhake, sahanpuru, nokiliette

ISBN (painettu) 978-952-60-8833-4**ISBN (pdf)** 978-952-60-8834-1**ISSN (painettu)** 1799-4934**ISSN (pdf)** 1799-4942**Julkaisupaikka** Helsinki**Painopaikka** Helsinki**Vuosi** 2019**Sivumäärä** 148**urn** <http://urn.fi/URN:ISBN:978-952-60-8834-1>

Acknowledgements

This doctoral thesis was carried out at Aalto University during the years 2010 – 2019. The financial support received from the following bodies is gratefully acknowledged: the Doctoral Program in Energy Efficiency and Systems (EES, funded by the Academy of Finland), Efficient Energy Use (EFEU) research program co-ordinated by CLEEN Ltd., Pohjolan voima Plc, Finnish Foundation for Technology Promotion, Foundation of Heikki and Hilma Honkanen, and Aalto University.

My deepest gratitude goes to my thesis advisor, Henrik Holmberg, D.Sc. Without your guidance and support, this dissertation would not exist. I would also like to thank my supervising professor, Pekka Ahtila, for his encouraging support. Let me also not forget to mention my former and present colleagues, Turo Haikonen, Timo Laukkanen, Mikko Kouhia, Aira Hast, Tuomas Paloposki and Ari Seppälä who created such a fun atmosphere at work. Professor Jonas Berghel from Karlstad University and Professor Margareta Björklund-Sänkiahö from Åbo Akademi are also greatly acknowledged for pre-examining this dissertation.

Finally, I express my deepest gratitude to Esko and Terhi, my parents, for their unconditional help and support. Tanja, thank you for your valuable friendship! The greatest of thanks to my husband, Valtteri, for being part of our family's fun life and for his support and love. I would also like to thank the sweetest little big man, Felix, for bringing indescribable joy into my life! In the end, I thank my little new baby boy, Max, for doubling my current happiness!

Espoo, Mäkkylä
16 October 2019
Tiina Myllymaa (née Järvinen)

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List of Abbreviations and Symbols

A_{dryer}	cross-sectional area of a continuous-working belt dryer [m^2]
AF	annuity payment factor [-]
A_{inv}	annual investment costs [€/a]
A_{oper}	annual operational costs [€/a]
a	specific volumetric heat transfer area / evaporation surface [m^2/m^3]
C_{CO2}	price of the CO ₂ emissions [€/t-CO ₂]
C_{elec}	price of the electricity [€/MWh]
C_{fuel}	price of the heavy fuel oil [€/t-oil]
C_{heat}	price of the heat [€/MWh]
$C_{sawdust}$	price of the sawdust [€/MWh]
$c_{p,da}$	specific heat capacity of dry air [kJ/kg°C]
$c_{p,w}$	specific heat capacity of water vapour [kJ/kg°C]
h	enthalpy of humid air [kJ/kg _{da}]
i	rate of interest [%]
$I_{tot,dryer}$	total investment costs of the dryer [€]
l_v	vaporization heat [J/kg]
m_0	drying rate during the constant drying period [kg _{H2O} /(kg _{db} s)]
$\dot{m}_{da}$	mass flow of dry air [kg _{da} /s]
$\dot{m}_{db}$	mass flow of material [kg _{db} /s]
$\dot{m}_{sawdust}$	mass flow rate of sawdust [kg _{db} /s]
$\dot{m}_{soot}$	mass flow rate of soot sludge [kg _{db} /s]
$\Delta \dot{m}_{fuel}$	reduced heavy fuel oil consumption [kg/h]

n	economic lifetime [a]
P	adjusted annual payment/drying costs [€/a]
P_{elec}	electricity consumption of the dryer fans [MW]
$P_{savings,CO_2}$	annual savings from reduced CO ₂ -emissions [€/a]
$P_{savings,fuel}$	annual savings in fuel consumption [€/a]
$P_{sawdust}$	annual costs of sawdust [€/a]
ΔP_{he+ad}	pressure drop over the plate heat exchanger(s) and the air duct system [Pa]
$\Delta P_{material}$	pressure drop over the material bed [Pa]
ΔP_{pres}	total pressure drop over the dryer configuration [Pa]
q_{oil}	heating value of heavy fuel oil [MJ/kg-oil]
$q_{sawdust}$	lower heating value of sawdust [MJ/kg _{db}]
r	mass ratio (dry basis) of sawdust to soot sludge in the mixture [-]
S_{CO_2}	CO ₂ emission factor of heavy fuel oil [kg-CO ₂ /MWh]
S_{net}	annual net savings [€/a]
T	air temperature [K]
t	air temperature [°C]
t_{wb}	wet bulb temperature [°C]
u	water content of material divided by weight of bone-dry material in the solid [kg _{H2O} /kg _{db}]
v	velocity of air before the fan(s) [m/s]
w	water content of material divided by total mass of wet material [kg _{H2O} /kg _{wb}]
X_{cr}	critical moisture content [kg _{H2O} /kg _{db}]
x	moisture content of air [kg _{H2O} /kg _{da}]
z	bed height of material inside the dryer [m]

α	heat transfer coefficient for the heat transfer between the bed and drying air in the boundary layer (convective) [W/m ² K]
δ_{da}	density of dry air [kg _{da} /m ³]
ρ_{db} ($= \rho_{bulk}$)	bulk density of dry material [kg _{db} /m ³]
τ	residence time of material inside the dryer [s]
γ	annual operating time of the dryer [h/a]
$\emptyset$	heat consumption of the dryer [MW]
η_{fan}	efficiency of the fan(s) [-]
BAT	Best Available Techniques
CDC	Characteristic Drying Curve
CHP	Combined Heat and Power plant
da	dry air
db	dry basis
DC	experimentally measured Drying Curve
LHV	lower heating value
MILP	Mixed Integer Linear Programming model
NPV	net present value
ORC	Organic Rankine Cycle
PBP	Payback Period
VOC	Volatile Organic Compound
vol-%	volume-%
wb	wet basis

Relation of the moisture contents:

$$w = \frac{u}{1+u}, \quad u = \frac{w}{1-w}$$

where $w = \frac{\text{moisture content (\%)}}{100 \%}$ is moisture content on wet basis [kg_{H2O}/kg_{tot}]
and u is moisture content on dry basis [kg_{H2O}/kg_{db}].

List of Publications

This doctoral dissertation consists of a summary and of the following original publications that are referred to in the text as I-IV. The publications are reproduced with permission from the publishers.

- I Myllymaa, Tiina; Holmberg, Henrik; Hillamo, Harri; Laajalehto, Tatu; Ahtila, Pekka. Wood Chip Drying in Fixed Beds: Drying Kinetics and Economics of Drying at a Municipal Combined Heat and Power Plant Site. *Drying Technology*, 2015, volume 33, issue 2, pages 205-215. Print ISSN of journal: 0737-3937. Online ISSN of journal: 1532-2300. DOI:10.1080/07373937.2014.945179.
- II Myllymaa, Tiina; Holmberg, Henrik; Arhippainen, Pentti; Ahtila, Pekka. Upgrading of Soot Sludge from Waste to Fuel by Means of Low Temperature Drying in Fixed Beds. *Fuel Processing Technology*, 2015, volume 138, pages 509-519. ISSN of journal: 0378-3820. DOI:10.1016/j.fuproc.2015.06.030.
- III Myllymaa, Tiina; Holmberg, Henrik; Ahtila, Pekka. Economic Evaluation of Drying of Soot Sludge and Sawdust Mixture at Low Temperatures Using the Characteristic Drying Curve Method. Submitted to Special issue of *ChemEngineering* (ISSN 2305-7084).
- IV Myllymaa, Tiina; Holmberg, Henrik; Ahtila, Pekka. Techno-Economic Evaluation of Biomass Drying in Moving Beds: The Effect of Drying Kinetics on Drying Costs. *Drying Technology*, 2019, volume 37, issue 10, pages 1201-1214. Print ISSN of journal: 0737-3937. Online ISSN of journal: 1532-2300. DOI:10.1080/07373937.2018.1492615.

Author's Contribution

Publication I: Wood Chip Drying in Fixed Beds: Drying Kinetics and Economics of Drying at a Municipal Combined Heat and Power Plant Site

Myllymaa is the corresponding author. The concept of the paper was planned by Myllymaa, Holmberg and Ahtila. Experimental tests were managed by Myllymaa, Hillamo and Holmberg. Myllymaa held the main responsibility of analysing the data and producing the calculations. Laajalehto developed an Excel-based calculation model for his research, which Myllymaa partly used in her research. The paper was written by Myllymaa. Holmberg provided valuable suggestions and comments.

Publication II: Upgrading of Soot Sludge from Waste to Fuel by Means of Low Temperature Drying in Fixed Beds

Myllymaa is the corresponding author. The concept of the paper was planned by all authors. Experimental tests were managed by Myllymaa and Holmberg. Myllymaa was responsible for analysing the data, producing the calculations and writing the paper. Holmberg provided valuable suggestions and comments.

Publication III: Economic Evaluation of Drying of Soot Sludge and Sawdust Mixture at Low Temperatures Using the Characteristic Drying Curve Method

Myllymaa is the corresponding author. The concept of the paper was planned by all authors. Myllymaa carried out the calculations and wrote the paper. Holmberg provided suggestions and comments.

Publication IV: Techno-Economic Evaluation of Biomass Drying in Moving Beds: The Effect of Drying Kinetics on Drying Costs

Myllymaa is the corresponding author. Holmberg, Myllymaa and Ahtila planned the concept of the paper. Myllymaa produced the model and calculations, and wrote the paper. Holmberg provided significant suggestions and comments.

1. Introduction

1.1 Background

Between the years 1900 and 2016, global primary energy consumption has increased from 12100 TWh to 151550 TWh in tandem with the economic growth and well-being of society [1]. In 2016, Finnish industry used energy 141 TWh, of which the largest consumers were the forest, metal and chemical industries [2]. Moreover, by 2017, the total emissions of green house gases in Finland were 56.1 milj. tonne CO₂-eq. The most significant sources of emissions were the energy sector (74 % of the total emissions), agriculture (12 %), and industrial processes (11 %). Within the energy sector, the most significant sources of emissions were the energy industry, domestic traffic, industry, and construction. [3] Most emissions from the energy sector result from the use of fossil fuels, which have accelerated climate change.

By 2030, the European Union hopes to achieve a 40 % cut in emissions from the level of the year 1990 in order to control climate change [4]. Additionally, by 2030, the share of renewable energy is expected to be 27 %, and energy efficiency to be improved by 27 % from the 1990 level [4]. The improvement in energy efficiency of processes is seen as one of the most promising and cost-effective measures for reducing global emissions and dependence on fossil fuels [5]. Improving the energy efficiency may have a significant impact on both primary energy consumption and CO₂ emissions.

The energy efficiency of industrial processes can be intensified by using more effective unit processes and process integration. Methods to improve industrial energy efficiency are for example different intensification methods of combustion, insulation, CHP production, heat storage, heat recovery, more advanced pumping and steam systems [6] as well as material recycling (e.g. in the pulp and paper industry) [7]. In [6], the Best Available Techniques (BAT) for improving energy efficiency in Finnish industry have been widely covered.

One opportunity to enhance energy efficiency is the utilization of excess heat (also known as waste heat) in order to decrease primary energy use. The energy used by the industry is usually divided into primary energy and secondary energy. In this dissertation, primary energy refers to fuel or electricity obtained outside the mill, or produced and used by the mill itself. Most primary energy converts into heat in the process in which primary energy is used. In this dissertation, the generated heat from the process is called secondary energy when it is utilized in some other process. Most of the primary energy exported to the processes is still used as secondary energy in other processes. The thermal

energy which is not used at the mill site or outside the mill is called excess or waste heat. The temperature level of the excess heat is typically between 40 – 100 °C [8], thus limiting its possible utilization. Primarily, new heat sinks for the excess heat should be found at the mill site. The most economic practice is to use the excess heat in a unit process where it is formed because utilization and formation are simultaneous and close to one another, in which case, the rate of utilization is high and investment costs are low. Nevertheless, it is also possible to utilize the excess heat outside the mill. Possible heat sinks include district heating [8] and the cultivation of algal biomass [9]. Upgrading lower heating values of moist fuels or waste streams by drying is also a potential heat sink for the excess heat [10].

Many industrial processes generate waste streams with a high moisture content and low heating value. These streams are usually disposed of by combusting them in boilers, and in this dissertation term waste stream always refers to a waste stream that can be combusted. The most common combustible waste streams are wood-based waste streams and different sludges. Solid wood-based wastes include woodchips, sawdust and bark [11]. The moisture content of solid wood-based biomass fluctuates generally from 40 to 60 % (wet basis) [12]. Other wood-based wastes are black liquor and sludges in the forest industry [11]. Sometimes the moisture content of the moist waste is excessively high (e.g. moisture content of soot sludge is 95 % wb) rendering it unable to combust. In this case, additional energy is needed if it is disposed of in a combustion chamber. It is also normal that some additional energy is needed in a combustion chamber despite the heating value of the waste being positive. Additional energy is needed to maintain a sufficiently high combustion temperature.

The lower heating value (LHV) of waste can be improved by drying the waste stream before combustion. In combustion process, fuel input, or alternatively live steam production, can be increased by up to 15 % due to the improved heating value [13]. Reduction in fuel consumption depends on the initial and final moisture contents of waste. For very moist fuels, improvements might be even greater. In many cases, drying is also a necessary unit process [14]. For example, when biodiesel is produced from wood, the biomass needs to be dried to a moisture content of 10 – 15 % (wb) before gasification [15]. While drying enhances the lower heating value, it also improves the quality of the biomass and its combustion. In addition, utilization of wastes as fuels after drying and the use of excess heat for heating the drying air enhance sustainable development.

There are several important factors that must be defined when a dryer is designed. These are drying gas (air, flue gas, steam), main heat transfer mode (convection, conduction, radiation), dryer type (such as conveyor/ belt dryer, drum dryer), the operational principle of a dryer (batch type or continuous working), and methods to improve the energy efficiency of the dryer (for example, heat recovery from the exhaust drying gas: using heat exchangers, recirculating exhaust gases). [12, 16] All the previously mentioned factors impact drying costs, which should be as low as possible.

A suitable and commonly used dryer type for low temperature drying is a conveyor/ belt dryer [17, 18, 19]. In [20], the range for the drying air temperature in a conveyor dryer is reported to be 30 – 150 °C. In this dissertation, a belt dryer working with air as drying gas is always assumed to be the dryer type when waste streams are dried. The benefits of a belt dryer include its simplicity and robust structure. Belt dryer can be used for wide range of materials and is therefore a commonly used dryer in the industry. However, low drying gas temperatures increase the dryer size and investment costs.

The main reason for this dissertation has been condensed in a direct citation from [12, p. 402]:

“As with any industrial dryer, the operating cost of a conveyor dryer is typically a significant contributor to the overall production cost of the finished product. A common mistake when purchasing a conveyor dryer is to underestimate the value of incremental improvements in dryer design. A small improvement in drying uniformity can often mean substantial savings over the life of the dryer.”

This dissertation shows results that support this statement well.

1.2 Objectives and structure of the dissertation

The total drying costs of the belt dryer may vary considerably depending on the decisions made for certain design parameters, such as bed height, air temperature and air velocity. Factors that also affect the drying costs include the heat source, its price, and its temperature level. The temperature level affects the amount of drying air, residence time of material in the dryer as well as the dryer size. The other affecting factor is, for example, the cost of the electricity. [12, 20]

The main objective of this dissertation is to investigate the effects of the design parameters of the belt dryer on total drying costs when different waste streams are dried using low temperature excess heat for heating the drying air. The influence is assessed by changing the following design parameters: bed height, drying air temperature, air velocity, and initial/final moisture contents of the material.

The waste streams included are woodchips, bark, as well as soot sludge and sawdust mixture. From the viewpoint of this study, the most interesting industrial waste stream is probably soot sludge, which is formed in the production process of formic acid and hydrogen peroxide. The moisture content of the sludge can be even 95 % wb. The soot sludge is normally disposed of by combusting. Heavy fuel oil is mixed with the soot sludge to be able to combust it in a boiler. Decreasing the moisture content of the sludge by drying, decreases the use of fuel oil.

The purpose of this dissertation is to find out optimal design parameters for the waste streams included in this work, and further, to consider which parameters have the most decisive effect on the drying costs; the profitability of drying is also evaluated during these investigations.

In order to produce a techno-economic evaluation for belt drying, the drying kinetics of the materials must be known; therefore, a series of experimental tests have been conducted for woodchips, bark, and the mixture of the soot sludge and sawdust to gain information on their drying kinetics in fixed beds. The soot sludge was mixed with sawdust to enable its drying in a belt dryer (for more information see page 29). Based on the data from the experiments, an Excel-based programme was developed to determine the drying costs. The dissertation is divided into the following sub-objectives:

- O1:** To determine drying kinetics in woodchip drying.
- O2:** To determine the drying kinetics in soot sludge and sawdust mixture drying, and to ascertain if the above-mentioned mixture can be dried at all.
- O3:** To form a generic model in order to determine the drying costs (annuity method) and profitability of drying when the energy contents of waste streams are upgraded by means of drying in a belt dryer.

This dissertation consists of four publications. Figure 1 summarises the scope. Drying data and some of the findings from Publications I, II, and III are used as a basis for Paper IV. In addition, data from an article concerning bark drying in fixed beds [21], which is not the author’s own, is used with permission in this dissertation. Data from Publ. II is used in Publ. III.

Figure 1. Connection of the Publications I – IV. Data from the article [21] concerning bark drying in fixed beds is not author’s own publication.

The structure of this dissertation is the following: Chapter 2 presents the rationale of the low temperature drying. Chapter 3 discusses the materials used

in the experiments, and presents the experimental test equipment that has been used in all experimental tests. Chapter 3 also summarises evaluation methods that have been used in economic calculations. Chapter 4 presents the main results of the work. Conclusions are drawn and recommendations presented in Chapter 5.

1.2.1 Literature review

Several studies can be found in literature that investigate the drying of different kinds of materials; indeed, an extensive review article of biomass drying was published by Stenström [22] surveying different kinds of biomass dryers. The paper focuses on the years 2000-2016. Chen et al. [23] and Tunçal and Uslu [24] extensively reviewed different dehydration technologies, including the removal of water without vaporization and thermal drying, for various municipal and industrial waste sludges. In [25], Bennamoun et al. reviewed fundamental information on wastewater sludge drying.

Drying of waste sludges in fixed bed has been studied in [26 – 39]. In [26], Li et al. analysed the shrinkage effect on mass transfer during convective fixed bed drying of sawdust and wastewater sludge mixtures. In the first part of the study, experimental tests were performed in a cross-flow convective dryer. Initial moisture contents were 85.5 % for the sludge and for sawdust 30 % (wb). The drying air temperature was 80 °C in the tests. In the second part, the authors developed a drying model in order to identify the internal diffusion coefficient and convective mass transfer coefficient from the experimental data. The results confirmed that sawdust addition was beneficial as the mass transfer efficiency between the air and material increased. The study also showed that neglecting shrinkage phenomenon resulted in an overestimation for the internal diffusion coefficient. Li et al. [27] investigated the drying of mixtures of sewage sludge and sawdust in a fixed bed, which included examining the influence of mixing step, sawdust and sludge ratio and drying temperature. The drying air temperature was 50, 80 and 110 °C in the tests. The addition of sawdust was shown to have a positive impact on the drying process with a mass ratio of 2:8 (db). In [28], Li et al. continued to study the convective drying behaviour of sawdust-sewage sludge mixtures in a fixed bed using the X-ray tomography. [29] discussed the optimal management of a residual sludge dewatering unit using mechanical dehydration and thermal drying; experimental tests were performed for organic (PVC latex), mineral (catalyst support) and biological residual sludges (activated sludge). The results indicated that in many cases it is desirable to mechanically eliminate the maximum amount of water from pure sludge. Vaxelaire and Puiggali [30] have simulated the drying of residual sludge in a belt dryer. Their research showed that a belt dryer was not well adapted to activated sludge because of a crust phenomenon where a dry layer appears quickly at the surface while the centre stays wet in a drying process. Nevertheless, for PVC industrial sludge, a belt dryer is more efficient and allows drying with reasonable residence times.

[31-36] evaluated convective drying of wastewater sludge. Léonard et al. [31] considered the influence of back mixing on the convective drying of residual

sludges in a fixed bed. Back mixing was simulated by adding dried sludge to the moist material before extrusion. The drying tests were done using air at temperature of 105 °C. Bennamoun et al. [32] performed experimental tests on two wastewater sludges in a discontinuous cross-flow convective dryer. They also studied shrinkage phenomena using mathematical modelling [32, 33]. Their research showed that the shrinkage effect must be taken into account when modeling a convective drying of the wastewater sludge and evaluating internal diffusion and convective mass transfer coefficients in wastewater drying. Forced convective drying of wastewater sludge was experimentally investigated and optimum drying conditions studied using exergy analysis in [34]. In [35], the influence of sludge origin in convective drying of five different wastewater sludges has been studied in which the results clearly indicated that sludge drying remains a particularly complex operation due to sludges possibly exhibiting very different behaviours, both from the kinetic and texture points of view. Fraikin et al. [36] used an experimental design on different wastewater sludges to highlight drying behaviour according to drying conditions, storage and extrusion. In [37], convective drying of sewage sludge was modelled using a reaction engineering approach. The research indicated that the reaction engineering approach can model the drying process well. Ali et al. [38] analysed solar convective drying of municipal sludge in thin layers. Huang et al. [39] assessed thermal behaviour of municipal sewage sludge in hot air drying (100 - 160 °C); the temperature profiles, moisture ratio and heat transfer coefficients of the thin layer sludge were obtained during drying in their research.

Several studies [21, 40-44] have investigated drying biomass in a belt dryers/ fixed bed where air is used as a drying gas. Holmberg et al. [21] experimentally investigated bark drying in fixed beds. In their research, normal drying curves and a characteristic drying curve were determined for bark. Four different drying air temperatures (50, 70, 90, and 110°C) and three bed heights (50, 150, and 250 mm) were used in the tests. Results indicated that the linear characteristic drying rate curve can be used for bark drying with good accuracy when the bed height is 150–250 mm and the inlet air temperature is below 100 °C. Bengtsson [40] experimentally analysed low-temperature (< 100 °C) bed drying of wooden biomass particles (woodchips and sawdust). The constant drying rate and falling drying rate drying periods were studied to obtain an energy efficient drying process. The tests showed that the drying zone progresses irregularly through the bed. Frodeson et al. [41] studied two-step drying techniques for improving energy efficiency and increasing drying capacity in fuel pellet production. In their study, sawdust (spruce, 52 % wb) was first dried in a packed moving bed dryer, and after that, in a pneumatic dryer. The average initial temperature of air in packed moving bed tests was 75 °C. When a pneumatic dryer was used, results showed that the drying capacity increased by 22 % as a result of the increase of the mass flow rate of sawdust. Huttunen et al. [42] presented a model which defines the capacity of a fixed bed bark dryer. In their study, a sensitivity analysis for different air temperatures, air velocities and bed heights showed that the dryer capacity (m³/h) depends on these parameters. Saastamoinen and Impola [43] modelled the drying of

biomass particles in fixed and moving beds with hot gas or steam. As a special case, they made experiments simulating woodchip drying in a fixed bed (air temperature 53 °C) and in counterflow. Their study showed that the size of the bed to reach a certain degree of drying depends mostly on particle size, particle moisture content, gas inlet temperature, gas inlet moisture content and gas mass flow rate. Pang and Xu [44] developed a mathematical model for a packed moving bed dryer of woody biomass (woodchips) as well as optimized the drying process. Results indicated that optimized drying process is achieved by using a countercurrent reversed flow, thus a lower air flow rate and lower drying temperature can be used to achieve the required moisture content. However, they did not validate the model with experimental tests.

Economic analyses of belt or fixed bed dryers used for biomass drying have been presented in [45–51]. Holmberg et al. [45] have used a MILP (Mixed Integer Linear Programming) model to minimize the drying costs of a continuous packed-bed bark dryer including both capital and operational costs. The MILP model was tested when three different heat sources: two warm waters (60 and 80 °C) and backpressure steam (0.4 MPa) were used. Their results indicated that the use of only low-temperature warm water flows for heating drying air is the most economic method when their prices are low (below 1 €/MWh). Li et al. [46] studied the integration of biomass drying into a power station fuel system evaluating its profitability using the net present value (NPV). Heat source (100 MW) for drying of pine woodchips were waste streams: superheated steam generated from the hot cooling water (90 °C) and flue gases (250–450 °C). According to this study, capital costs become higher when superheated steam is used. For both heat sources, a payback period of 3–4 years was calculated (interest rate 0 %), but the profitability was sensitive to fuel prices. Holmberg and Ahtila [47] compared the drying costs in biofuel drying (regularly shaped spruce particles) between multi-stage and single-stage drying with multi-stage heating in a pulp and paper mill. Secondary heat (78 °C), backpressure steam (4 bar, 138°C) and extraction steam (10 bar, 180°C) were the heat sources for drying. According to their results, single-stage drying is usually a more economic drying system when the amortisation time is short. Holmberg and Ahtila [48] optimized a soft wood bark drying process in a combined heat and power production of pulp and paper mill using the net present value. In [48], the used heat sources for drying have been secondary heat (70 °C), backpressure steam (120 °C) and extraction steam (150°C). Results indicated that using energy prices from the year 2005, and the economic lifetime of 10 years with an interest rate of 5 %, the optimally designed dryer used only secondary heat for the heating of air, consisted of two drying stages, and dried the bark flow to a final moisture of 0.3 kg/kg_{db}. Myllymaa et al. [49] studied the profitability of woodchip drying in a batch-type dryer at a municipal combined heat and power plant site using the payback period method. Gigler et al. [50] applied cost evaluations to willow chip drying with ambient air in a fixed deep bed using farm facilities. Furthermore, Le Lostec et al. [51] performed the thermal and economic analysis of woodchip drying in a conveyor belt with an absorption heat pump. In their research, three processes were studied: drying

wood with a heat pump powered by a wood burning furnace, direct heating of the air by the furnace and heating of the air using industrial waste heat. Their results indicated that of the three processes that were studied, industrial heat recovery systems proved to be the most energy efficient and economic solution.

Although the drying of waste streams has been extensively studied, previous studies did not systematically analyse the ways in which the most important initial drying parameters such as bed height, air velocity, temperature affected the drying costs of a belt drying.

2. Rationale

2.1 Availability of industrial excess heat in Finland

In Finland, industry is the most significant end user of energy, thus rendering it the sector with the most potential to improve energy efficiency in society [8, 52]. Indeed, significant amounts of excess heat are available in industry [8]. The excess heat is usually bound to flue gases, coolant waters, outgoing air, waste waters, exhaust steam, exhaust gases from dryers, process gases or condensation heat [8]. Improvement in energy utilization is fulfilled when less energy is needed to produce the same output, or when the same energy input is used but more services are delivered with it. Utilisation of excess heat can be described as a green and carbon neutral energy source as it is making use of what is essentially a waste product [53]. Figure 2 shows the difference between the terms primary energy, secondary energy and excess heat.

Figure 2. Schematic picture of the difference between primary energy, secondary heat and excess heat. (Established from [8]).

A basic requirement for the utilization of excess heat is that the energy recovered from excess heat (used as a heat or converted into mechanical work) must have an end-user.; otherwise, excess heat recovery is unavailing. The type (steam, warm/hot water, flue gas, warm air) and the temperature level of excess heat vary greatly. From the viewpoint of drying, it is important to know the source of the excess heat, its temperature, the amount of available excess heat and possible variations in its availability. All these factors affect if the excess heat can be utilized in a specific drying process.

Tables 1 and 3 – 6 show the excess heat potential of the three greatest industrial energy users: forest, chemical and metal industries in Finland. The manufacturing of timber, wood, cork, paper and board products are included in the forest industry, whereas the manufacturing of coke, oil refining and chemical products are incorporated into the chemical industry. The field of metal industry consists only of the sector called *refining of metals*. The excess heat potentials have been evaluated in the YIT report [54] by collecting data from specific mills in each sector with their calculated excess heat potential being then proportioned to the whole industrial sector in Finland. Tables 1 and 3 present the potential of the forest industry. Table 1 presents the excess heat potential and temperature ranges of various excess heat streams for the manufacturing of timber, wood and cork products. Table 2 shows different paper and board industry mills in Finland, which were included in the YIT report, and the excess heat potential of those mills. The original report of YIT [54] does not reveal precisely the kinds of mills they are, but shows examples for the temperature levels of the excess heat streams in the paper and board sector. Based on the excess heat potential of the examined mills, the potential for the whole sector was calculated in the YIT report. The excess heat potential of the whole paper and board sector is shown in Table 3. However, the data of

Table 3 does not include the excess heat of which the temperature is under 35 °C [54].

Table 1. The excess heat potential for the manufacturing of timber, wood and cork products [54].

Stream	Temperature [°C]	Excess heat potential of the sector [GWh/a]
Exhaust gas	120-155 / 135	1,938
Exhaust gas	65-25 / 55	2,626

Table 2. The excess heat potential of examined mills in [54].

Mill type	Stream	Temperature [°C]	Excess heat potential of the examined mill [GWh/a]
Mill A1	flue gas	126	81.8
	air	50-70	49.2
	water	41.5	3.2
Mill A2	air	40-50	16.2
	water	43-45.2	24.6
Mill A3	air	-	25.9
	water	39-41	10.6
Mill B4	flue gas	-	108.8
	air	50-70	185.8
	water	70-75	147.2
	water	70	56
Mill B5	air	50-70	222.3
	water	42-52	79.4
	steam	100-120	168
Mill B6	air	50-70	135.7
	water	40-45	131.5
	steam/ water	100 / 75	118.4
Mill C7	air	50-65	203.8
	water	80	39.3
	water	65-70	44.7
	water	70	96.8
	flue gas	160	24
	flue gas	160	347.5
Mill C8	air	60-70	203.8
	water	80	249.5
	water	46	23.1
	water	65-70	85.6
	flue gas	180	62.4
	flue gas	180	451.8
Mill C9	air	60-70	78.4
	water	70-80	20.8
	air/ steam	60 / 140	15.2
	flue gas	120	55.2
TOTAL:			3,566.5

Table 3. The excess heat potential for the whole sector of manufacturing of paper, board and paper products [54].

Temperature level	35 – 55 °C	55 – 120 °C
	Excess heat potential [GWh/a]	Excess heat potential [GWh/a]
Mill group A	773	147
Mill group B	3331	1140
Mill group C	7749	5600
Sum	11853	6887

Tables 4 and 5 present the excess heat potential of the chemical industry. The potential of the manufacturing of coke and oil refining products, and the manufacturing of chemicals and chemical products are shown in Tables 4 and 5, respectively. It should be noticed that the excess heat potential of the manufacturing of chemicals and chemical products sector is evaluated only for one Finnish mill called Borealis Polymers Oy in the YIT report. The excess heat potential of the metal industry is shown in Table 6.

Table 4. The excess heat potential for the manufacturing of coke and oil refining products [54].

Stream	Temperature [°C]	Excess heat potential of the sector [GWh/a]
Water	20-40	7,525
Flue gas	170-600	2,167
Air	40-190	2,296

Table 5. The excess heat potential for the manufacturing of chemicals and chemical products [54].

Stream	Temperature [°C]	Excess heat potential of the sector [GWh/a]
Water	>70	360
Water	80-90	240
Water	65-70	240
Water	-	2,160

Table 6. The excess heat potential for the refining of metals [54].

Stream	Temperature [°C]	Excess heat potential of the sector [GWh/a]
Water	over 50	1,830
Flue gas/ process gas	250-700	4,800
-	below 55	8,270

Tables 1 and 3- 6 shows that the total excess heat potential of the three greatest industries is 53192 GWh/a. When the total excess heat potential is changed to the total theoretical exergy potential, the value is 6.9 TWh/a [55], which is 10 % of the electricity produced in Finland in 2011. The total theoretical exergy potential describes the maximal amount of mechanical work which could be produced from the excess heat, for example by using organic Rankine cycle (ORC) or Kalina cycle [56].

Table 7 summarises the amount of excess heat available at different temperature levels in the three greatest industrial sectors in Finland [55]. The highest temperatures are in the refining of metals and chemical industry: 75 % and 52 % of the theoretical exergy potential of the field are at temperatures over 200 °C. In the forest industry, most of the theoretical exergy potential is at the temperature range of 100 - 200 °C.

Table 7. The relative share of the available exergy potential at different temperature levels for each field [55].

Industrial field	T ≤ 55 °C [%]	55 °C < T ≤ 100 °C [%]	100 °C < T ≤ 200 °C [%]	T > 200 °C [%]
Forest industry	9	32	60	0
Chemical industry	20	5	22	52
Refining of metals	25	0	0	75

Nevertheless, the excess heat recovery represents a great opportunity to enhance the energy efficiency of systems and mitigate climate change. Although heat from the processes is already partly exploited in industry, the huge potential for enhancement still exists. For example, the use of secondary heat of industry in district heating in Finland has increased from 775 GWh (2010) to 1421 GWh (2015) [57]. More than a third of the energy used in the Finnish industry turned into excess heat without heat recovery and utilization in 2010 [58].

2.2 Drying

2.2.1 Benefits of drying

A significant amount of water moves along waste streams through the system boundaries of a power plant. When moist materials are combusted, the water of the material vaporises inside the boiler. Energy used for vaporising the water from the material decreases the energy amount used to produce high-pressure steam. Lower steam production in a boiler results in lower power and heat production at a CHP plant. [59]

The most significant benefit of drying before combustion is the improved lower heating value of fuel (also known as effective heating value). The lower heating value is directly proportional to the moisture content of the material [60]. The same amount of dry matter produces more energy, or less dry matter is needed in a boiler to produce the same amount of steam. The fuel input depends on the power and heat production at a CHP plant. Usually, the fuel input is desired to keep as a constant and due to that, drying decreases the use of a marginal fuel in a boiler. The term ‘marginal fuel’ means fuel which is more expensive at the time depending on the fuel price and the price of CO₂-tonne.

Drying also decreases the variation of the quality of the biomass, which is useful from the viewpoint of the boiler operation. [14] Drying also decreases dry matter losses, if the biomass is stored before the combustion. Moisture of biomass causes, for example, mould, composting of the organic matter and a fall of needles, which all cause the dry matter losses in storing [61]. This may be particularly beneficial for municipal CHP plants that combust biomass. Municipal CHP plants are not in operation during the summer, but sometimes biomass is stored on the mill site during the summer months.

If biomass is dried outside the plant where it is used, drying also decreases transportation costs as it reduces the mass of the biomass due to water vaporisation.

The lower moisture content of the biomass increases the adiabatic combustion temperature and decreases the necessary residence time in a combustion chamber [62]. The higher moisture content would result in incomplete combustion, which increases the emissions [62, 63]. For example, increasing the moisture content means higher emissions of hydrocarbons [63, 61]. Incomplete combustion also causes higher particle emissions [61].

Raitila et al. [64], Liang et al. [65], Koskiniemi and Viirimäki [66], and Flyktman and Helynen [67] have also reported that the boiler efficiency improves when dried wood-based biomass is combusted instead of the moist one. However, the biomass drying can cause the boiler some challenges. As the lower heating value increases via drying, the adiabatic combustion temperature simultaneously increases. This can lead to agglomeration and sintering in a bubbling fluidized boiler.

2.2.2 Low temperature drying – Belt dryer

Dryers can be classified according to the drying media: air, superheated steam, flue gas or vacuum. Dryers using flue gas or air as the drying gas are usually used for biomass drying. Dryers can be also divided into groups on the basis of the main heat transfer mode: convective (direct) dryers, conductive (indirect) dryers and dryers where heat transfer is based on radiation. Dryers in which heat from the drying gas is brought to the material by convection are more common in biomass drying. Different dryer types include rotary dryer, fluidized bed dryer, belt/conveyor dryer, pneumatic dryer, and spray dryer. Dryers can be also divided by their working method: continuous or batch dryer.

Dryers with low-temperature drying gas are usually belt dryers (also known as conveyor dryers) in which heat transfer is based on convection [17-19]. In practice, the most common drying gas for low-temperature drying is air.

The most notable emissions occurring in biomass drying are particle, Volatile Organic Compound (VOC) and odour emissions. [12, p. 858] VOCs may contribute air pollution and eutrophication. Most of the VOC emissions generated during wood drying are terpenes. VOC emissions react chemically with nitrogen oxides in the presence of sunlight forming ground-level ozone, which is harmful to the environment. [68] Most of the terpenes are released in the beginning of the drying [69]. However, using a low drying temperature (below 100 °C) decreases the amount of released emissions from the dryer [70, 71]. When the temperature is below 100 °C, emissions are minor and can be compared to natural emissions, which are naturally released in forests and storage, because thermal degradation is low [72]. Currently, there are no emission limits for biomass drying in which drying air and convection heat transfer are used [72]. Emissions from a belt dryer are discussed in more detail in [63, 70, 71, 73].

The belt dryer is also known as a packed moving bed dryer. It is one of the most versatile dryers capable of handling a wide range of products, for example, animal feed, nuts, and woody biomasses [12]. Belt dryers are used for several biomasses including woodchips, bark, forest residue and sawdust [40, 74]. Figure 3 illustrates a typical belt dryer.

Figure 3. A typical belt dryer [IV].

The structure of a belt dryer is quite simple and robust. The moist material is spread evenly by a feeder at constant mass flow rate on the conveyor. The drying air is forced by fans through the material bed and the conveyor, which is made of perforated sheet metal plates. [12] Outdoor air is heated in a heat exchanger before it passes the bed and the conveyor. A pressure drop is formed over the material bed on the belt dryer.

The drying air can flow through the material bed from top to bottom or vice versa. This air heats the material, vaporises the moisture and absorbs the vaporized moisture emerging as humid air from the dryer. Controlling the speed of the conveyor adjusts the residence time of material inside the dryer. Thus, the desired final moisture content of the material is achieved. Usually, the heating unit (Fig. 3) is an indirect heat recovery system in industrial processes. In the heat recovery system, heat from the heat source is first transferred into a circulating fluid, which then heats the drying air in a following heat exchanger [73]. The belt dryer is usually divided into multiple drying blocks (see Fig. 3). These zones have their own heating unit and fan. In each block, the drying air temperature, velocity or humidity may be different. [IV]

Due to the residence time of the material being easy to adjust to a desired one, the belt dryer is suitable for drying in which the residence time is long as in low-temperature drying. However, the surface area of the dryer is often large, especially if drying air at a low temperature is used, which increases the investment costs.

Various belt dryer configurations exist in which the conveyor and airflow arrangement differs [12]. The simplest belt dryer configuration is a single-stage belt dryer in which there is one conveyor (Fig. 3) [73]. In multiple-conveyor dryers, there are several stages in the series, or the conveyors can be situated above each other running in opposite directions, which reduces the required floor space for the dryer. All the stages can run on different speeds and bed heights, which enhances the drying process. [12, 73]

2.2.3 Total drying costs

Total drying costs take into account both the investment and operational costs of the dryer. The following main factors affect the total drying costs of the belt dryer [12, 20, 42, 45, 46, 47]:

- Cost of the heat
- Cost of the electricity
- The dryer size
- Annual operating time of a dryer

The cost of the heat energy usually depends on the height of the temperature of the heat source (backpressure steam versus excess heat). If the heat is priced correctly, the price always depends on the temperature level of the heat source: the higher the temperature, the higher the price. However, in terms of the real world, the price is not always dependent on the temperature level. [IV]

The pressure drop over the material bed has an influence on the electricity consumption of the air fan, and consequently, on the electricity demand of the whole drying system. Even if the heat source for heating the drying air would be free of charge (for example, excess heat), the volume flow rate of air may become large, increasing the electricity consumption and operational costs of the dryer. [I]

The design parameters of a belt dryer that affect the dryer size and the investment costs are the following: initial and final moisture contents of the waste material, drying air temperature, air velocity, bed height, waste material which is needed to dry, mass flow rate of waste material inside the dryer, and bulk density of the material.

Above-mentioned factors are taken into account in this dissertation. In addition, personnel costs [46, 47], maintenance costs [46, 47], insurance [46, 47], pumping costs [47], rate of interest and utilization time of a dryer have influence on drying costs. On the other hand, the possible existing and utilizable infrastructure that can be used for a dryer may reduce the total dryer investment costs [72].

3. Materials and methods

3.1 Materials

3.1.1 Woodchips

Wood-based fuels are typically divided into three main groups: black liquor, sludge, and solid wood fuels. The most important of solid wood fuels are bark, sawdust, cutter shavings, recycled construction wood, and woodchips. The term “woodchip” is a general item for whole tree chips or chips made from stem wood or logging residue, regardless of chipping place. [11]

The use of woodchips in Finnish energy production (heating plants and CHPs) has increased from 0.8 to 7.2 million solid cubic meters between the years 2000 and 2017. When measured by volume, woodchips are the second most significant wood fuel after bark in Finland. [75] The cost of woodchips has fluctuated in Finland between 20 €/MWh to almost 22 €/MWh between the years 2015 - 2018 [76].

The most important parameters affecting the quality of woodchips are moisture content, energy density, and particle size distribution [77]. Woodchips typically have an irregular particle shape and heterogeneous particle size distribution. The moisture content of woodchips typically varies between 40 and 65 % (wb).

The woodchips (Fig. 4) used in the experimental tests were obtained from a Finnish municipal CHP plant. Woodchips were mostly logging residue, which included bark, stem wood, and branches. The size variation of the chips was large, making the chips very inhomogeneous. The initial moisture content of the woodchips varied from 1.27 to 1.91 kg_{H₂O}/kg_{db} in the tests.

Figure 4. Woodchips used in the experimental tests. [1]

3.1.2 Bark

In the wood processing industry, bark is generated in the debarking process [12]. It is mostly obtained as a by-product from the pulp and paper mills in which it is usually combusted in the combined heat and power plants of the mills to produce heat and electricity. The moisture content of the bark after debarking varies typically between 50 – 60 % (wb), depending on the wood species, weather, and season. [21]

At present, bark is the most significant solid wood-based fuel used in Finland [75]. The cost of bark has fluctuated in Finland between 15.5 €/MWh to 16.7 €/MWh between the years 2015 - 2018 [78]. Figure 5 shows the bark used in the experimental tests of Holmberg et al. [21]. The bark has been obtained from a Finnish pulp mill after the dry debarking. The initial moisture content of the birch bark varied from 0.99 to 1.52 kg_{H₂O}/kg_{db} in the tests.

Figure 5. Birch bark used in the experimental tests. [21]

3.1.3 Soot sludge and sawdust

The soot sludge used in this dissertation is a waste stream, which is formed in the production process of formic acid and hydrogen peroxide at the production plant of Eastman Chemical Company, situated in Oulu, Finland [79]. Soot sludge has an extremely high moisture content (95 % wb). In addition, soot sludge has a very sticky and viscous nature. Currently, moist soot sludge is disposed of by combustion. Fuel oil is mixed with soot sludge in order to combust it. Oil mixing also reduces the viscosity of soot sludge, thus enhancing its pumping properties. The theoretical lower heating value of the moist sludge is $-23.8 \text{ MJ/kg}_{\text{db}}$ which is obtained as follows:

$$\text{LHV} = \text{LHV}_{\text{dry, soot sludge}} - 2.443 * u_{\text{soot sludge}} = 22.65 \text{ MJ/kg}_{\text{db}} - 2.443 \text{ MJ/kg}_{\text{H}_2\text{O}} * 19.0 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}.$$

Figure 6 presents the formation process of the soot sludge.

Figure 6. Formation process of the soot sludge. [III] (Reproduced from [80].)

Among other products, the Eastman Chemical Company produces formic acid and hydrogen peroxide using heavy fuel oil as the main raw material. The production chain starts by gasifying the heavy fuel oil with oxygen (partial oxidation) and steam at high pressure to produce syngas. The syngas is mainly composed of hydrogen and carbon monoxide as well as carbon soot (1 – 2 % of the amount of used heavy oil) after gasification. After cooling in a waste heat boiler, the syngas is scrubbed with water, and the gas (CO , H_2 , CO_2 and H_2S) and soot water are separated. The gas is directed to an amine-wash for the recovery of sulphur and CO_2 , and then further to the separation of CO and H_2 , which are used for the production of formic acid and hydrogen peroxide. Soot sludge from the centrifugation process is mixed with oil before combustion to improve the flowability of sludge. Fuel oil is also used as a supporting fuel at a CHP plant. [80] In Eastman's process, moist soot sludge (c. 95 % wb) is formed approximately 20 t_{wb}/d. More detail information on the soot sludge (e.g. ultimate analysis) can be found from Publ. II.

Due to the extremely sticky, moist, viscous and homogenous nature of soot sludge (slightly similar to blacking), air cannot evenly flow through the bed if the bed is only soot sludge. In order to get an even airflow through the entire bed, the sludge was mixed with sawdust in the experimental tests. In Publication

II, sawdust was chosen as a mixing material due to its availability at the production site. This mixing considerably increased air channels in the bed, thus enhancing the drying process. Li et al. [26] studied the convective bed drying of extruded wastewater sludge and sawdust mixture. They found that when having reached a certain mixture ratio, the drying time decreases compared to drying with only the sludge. This observation also supports drying of soot sludge together with sawdust. Drying soot sludge converts the negative heating value into a positive one, and decreases oil use.

Generally, sawdust is formed in timber sawing in the wood processing industry. The typical moisture content of moist sawdust is 40 – 60 % (wb). [12] The price of the sawdust has varied from 14 to 16 €/MWh between the years 2015 to 2018 [78]. Figure 7 shows examples of the soot sludge and sawdust used in the experimental tests.

Figure 7. Moist soot sludge (approx. 95 % wb) (a), and moist sawdust (approx. 56 % wb) (b). [11]

In the tests, the moisture contents of the soot sludge and sawdust ranged from 94.9 to 95.5 % (wb), and from 53.9 to 58.1 %, respectively. Three mixture ratios (vol.-%) for the soot sludge and sawdust were used in the tests (30%_{soot}: 70%_{sawdust}, 50%_{soot}: 50%_{sawdust} and 70%_{soot}: 30%_{sawdust}). The range of the moisture contents for various mixture ratios (vol.-%) were the following:

- 30%_{soot}: 70%_{sawdust}: 76.8–78.1 % (wb),
- 50%_{soot}: 50%_{sawdust}: 83.7– 85.4 % and
- 70%_{soot}: 30%_{sawdust}: 90.1 %.

The corresponding mass ratios (% wb, $m_{\text{soot,tot}}/m_{\text{mixture,tot}}$) of the mixtures were the following:

- For volume ratio of 30%_{soot}: 70%_{sawdust}:
Mass ratio 55–58 %_{soot}, 42–45 %_{sawdust}
- Volume ratio of 50%_{soot}: 50%_{sawdust}:
Mass ratio 71–73 %_{soot}, 27–29 %_{sawdust} and
- Volume ratio of 70%_{soot}: 30%_{sawdust}:
Mass ratio 86 %_{soot}, 14 %_{sawdust}.

The mixture ratio of 50%_{soot}: 50%_{sawdust} was found to be the most suitable mixture ratio on the basis of the results in Publication II, and it was used to study the influence of air velocity, bed height and air temperature on the drying kinetics. Figure 8 shows an example of the moist soot sludge and sawdust mixture when the mixture ratio is 50 %_{soot sludge} and 50 %_{sawdust} (vol.-%).

Figure 8. Moist soot sludge and sawdust mixture (mixture ratio in volume-%: 50 % soot sludge and 50 % sawdust). [III]

3.2 Experimental tests – The drying equipment

The drying data used in all Publications I – IV has been produced using the same experimental test equipment shown in Figure 9. The tests were conducted in a fixed-bed batch-type dryer, where heat convectively transfers to the material. Pressurized air from the laboratory was first heated in an electric heater to a higher temperature than the desired drying air inside the dryer. Indoor hall air was then mixed with the pressurized and heated air in order to obtain the desired drying air temperature. By changing the rotation speed of the fan, the desired air velocity before the material bed was obtained. The drying air flowed through the material bed from top to bottom. The drying of the material was examined using a direct method. The drying chamber laid on a scale, thus saving the mass of the material bed on the computer after a 10-second time interval. The mass of the material bed decreased as the drying process proceeded. The drying curve for the material was then defined on the basis of the mass change and the initial moisture content.

The average initial moisture content of woodchips, soot sludge and sawdust (Tables 8 and 10) was defined by drying two to three material samples (mass of

a moist sample ~150–350 g depending on the material) in an oven at a temperature of 105 ± 2 °C for at least 16 hours, but not over 24 h [81]. The samples were taken from the same container in which the moist material for drying tests were stored. After the oven drying, the samples were absolutely dry and the initial moisture content could be defined on the basis of the mass change.

The drying air used in the tests was the indoor air of the laboratory hall. The average inlet air temperature before the bed varied ± 1 °C from the default temperature value in all the tests.

Before and after each test, the mass of the material samples was also measured using a separate scale in order to evaluate the accuracy of the test rig in measuring the mass change of the sample. The accuracy was evaluated by calculating the relative difference between the two mass measurements (Eq. 1).

$$\text{Relative difference} = \left| \frac{\Delta m_{\text{Test rig}}}{\Delta m_{\text{Separate scale}}} - 1 \right| \quad (1)$$

The absolute humidity of the inlet air was the same as the ambient air humidity inside the laboratory hall. The absolute humidity was defined by measuring the relative humidity and inlet air temperature using a humidity transmitter. [21]

Figure 9. a) The test rig at the laboratory at Aalto University. b) Flow diagram of the equipment. (The most important technical data of the equipment: maximum power of the heating unit 30 kW, maximum temperature of the drying air ~ 200 °C, maximum air velocity per free cross-sectional

area of the drying chamber ~1–1.2 m/s, height of the drying chamber 800 mm and diameter 400 mm.) [11]

3.2.1 Determination of drying curves

The drying curves used in the economic calculations are based on experimental drying data. All the residence times of materials in the dryer have been defined using these curves.

The drying curves of the woodchips and the mix of the soot sludge and sawdust were produced by the author of this dissertation. Those drying kinetics results are presented in Sections 4.1 and 4.2. The drying curves of bark, which are also used in this dissertation, can be found from the reference [21, page 956: Figures 3 and 4, p. 957: Figures 5 and 6].

The particle size distribution of the waste streams used in the tests might vary considerably (see e.g. Fig. 4), thus influencing the shape of the drying curves; therefore, two or three samples were dried using the same initial values in the tests. The raw drying data was handled according to the guidelines for processing experimental drying kinetics data presented in [82]. The actual drying curves were defined from the measurement data by determining a cubic polynomial regression fit (the least squares 3rd degree polynomial) of the experimentally measured drying curves with identical drying conditions.

Publications I and II studied the effects of different parameters (e.g. air velocities and bed heights) on drying time. This influence of different parameters was studied because the cross-sectional area of the dryer is directly proportional to the drying time.

3.3 Calculations

3.3.1 Determination of drying costs

The total drying costs of a dryer have been determined using the annuity method (Equation 2) in Publ. IV.

$$P = A_{inv} + A_{oper} = AF \cdot I_{tot,dryer} + A_{oper} , \quad (2)$$

in which P is adjusted annual payment/costs [€/a], A_{inv} is annual investment costs [€/a], A_{oper} denotes annual operational costs [€/a] and $I_{tot,dryer}$ is total investment costs of the dryer [€]. The term AF is the annuity payment factor:

$$AF = \frac{\frac{i}{100}}{1 - \left(1 + \frac{i}{100}\right)^{-n}} , \quad (3)$$

where i is the rate of interest [%] and n is the economic lifetime [a]. The residual value of a dryer is assumed to be zero in all cost calculations. The total investment costs are calculated using the following correlation:

$$I_{tot,dryer} = A_{dryer} \cdot \left[-3095 \ln \left(\frac{A_{dryer}}{480} \right) + 5838 \right] , \quad (4)$$

where A_{dryer} is the cross-sectional area of a continuous-working belt dryer [m²]. Equation 4 is based on investment costs obtained from current dryer manufacturers (Stela, Andritz, Swiss Combi). Investment costs per cross-sectional area of a dryer are shown in Fig. 10 [83, 84, 85]. Eq. 4 includes the cost of the dryer itself and the cost of heat exchangers and fans. It should be considered that Eq. 4 is accurate when $0 \text{ m}^2 < A_{dryer} \leq 480 \text{ m}^2$. This is because the largest dryer that can be produced by one of the component manufacturers, included in this study, is 480 m².

Figure 10. Left graph: the dryer investment cost data from the dryer manufacturers. Right graph: the dimensionless equation for the dryer investment cost. [IV]

A_{dryer} can be approximately determined as follows [45]:

$$A_{dryer} = \frac{\dot{m}_{db} \cdot \tau}{\rho_{bulk} \cdot z}, \quad (5)$$

in which $\dot{m}_{db}$ is the mass flow of dry material [kg_{db}/s], τ is the residence time of material inside the dryer [s], ρ_{bulk} represents the bulk density of dry material [kg_{db}/m³] and z is the bed height of material inside the dryer [m]. The residence time (τ) in Eq. 5 is equal to the drying time of material which has been dried experimentally in a batch-type fixed bed dryer. The residence times (τ) of materials in the dryer are based on experimentally measured drying curves/data published in [I, II, 21]. All experiments were conducted with the same test rig (see Fig. 9).

The annual operational cost (A_{oper}) in Eq. 2 is expressed as follows:

$$A_{oper} = P_{elec} \cdot C_{elec} \cdot \gamma + \emptyset \cdot C_{heat} \cdot \gamma, \quad (6)$$

where γ is the annual operating time of the dryer [h/a], P_{elec} denotes the electricity consumption of the dryer fans [MW], $\emptyset$ represents the heat consumption of the dryer (heating of drying air) [MW], C_{elec} is the price of the electricity and C_{heat} is the price of the heat [€/MWh]. The electricity consumption is calculated as follows:

$$P_{elec} = \frac{v \cdot A_{dryer} \cdot \Delta P_{pres}}{\eta_{fan}}, \quad (7)$$

where v is the velocity of air before the fan(s) [m/s], ΔP_{pres} represents the total pressure drop over the dryer configuration [Pa] and η_{fan} is the efficiency of the fan(s). It is assumed that the fan(s) are situated before the heat exchanger(s) in the dryer configuration. The total pressure drop ΔP_{pres} is the sum of the pressure drops over the material bed, the plate heat exchanger(s) and the air duct system (i.e. $\Delta P_{pres} = \Delta P_{material} + \Delta P_{heat+ad}$). Equation 8 calculates the heat consumption of the dryer for Eq. 6.

$$\dot{Q} = \dot{m}_{da} \cdot (h_{out} - h_{in}) = v \cdot A_{dryer} \cdot \delta_{da} \cdot (h_{out} - h_{in}) \quad (8)$$

in which $\dot{m}_{da}$ is the mass flow rate of dry air [kg_{da}/s], h denotes the enthalpy of humid air before and after the heat exchanger [kJ/kg_{da}] and δ_{da} is the density of dry air [kg_{da}/m³] at the drying air temperature. The enthalpy of humid air is calculated using Equation 9.

$$h = c_{p,da} \cdot t + x(c_{p,w} \cdot t + 2501) \quad (9)$$

where $c_{p,da}$ is the specific heat capacity of dry air [kJ/kg°C], t denotes the temperature of air [°C], x is the moisture content of air [kg/kg_{da}] and $c_{p,w}$ is the specific heat capacity of water vapour.

The main results of this dissertation (Section 4.3) have been calculated using an Excel-based simulation programme created by Myllymaa. The programme calculates the total drying costs (Eq. 2) for a specific drying combination.

3.3.2 Characteristic drying curves

Usually, drying times or rates are needed in order to design a dryer. One option is to use drying models to define the drying time. Unfortunately, drying models may sometimes yield extremely inaccurate estimates of the real drying time. Experimental drying tests usually evaluate drying times more precisely. However, large amounts of measurements are needed to define drying curves under various drying conditions. Experimental drying tests are usually time-consuming and they can even be costly. A single drying curve, called a characteristic drying curve (CDC), can be formed by normalizing measured drying curves with respect to the constant drying rate and critical moisture content. A characteristic drying curve can be used to extrapolate drying data from one set of external conditions to another. This reduces the number of measurements. Furthermore, a single curve also describes the drying rate over a wide range of different drying conditions. [21, 86]

In Publication III, characteristic drying curves were determined for a soot sludge and sawdust mix, which has been dried in fixed beds at different bed heights and air temperatures. To be able to define the constant drying rate in various drying conditions, a volumetric heat transfer coefficient was also defined for the mixture. In addition, characteristic drying curves were used in an economic analysis in which the profitability was evaluated of the soot sludge drying before combustion.

Drying phenomenon can be divided into three periods: a short initial heating period, constant drying rate period and falling drying rate period. The moisture content, after which the falling drying rate period begins, is called the critical moisture content. The moisture content, after which drying ends totally, is called the equilibrium moisture content. The equilibrium moisture content for the material can be read from the sorption isotherm.

The characteristic drying curve corresponds to the relative drying rate as a function of the relative moisture content. The relative drying rate (m_R) is defined as follows:

$$m_R = \frac{m}{m_0}, \quad (10)$$

where m is the drying rate at the time t and m_0 is the drying rate during the constant drying rate period. The relative moisture content (X_R) is defined as follows:

$$X_R = \frac{X - X_{eq}}{X_{cr} - X_{eq}}, \quad (11)$$

where X is the moisture content at the time t , X_{eq} represents the equilibrium moisture content, and X_{cr} denotes the critical moisture content. In many cases, X_{eq} approaches 0 and Eq. 11 results in:

$$X_R = \frac{X}{X_{cr}}. \quad (12)$$

When the characteristic drying curve is used, the drying rate is expressed as follows [86]:

$$m = m_0 f(X_R), \quad (13)$$

where m_0 is the drying rate in period of constant drying rate and $f(X_R)$ is the characteristic drying curve. The change of the moisture content of the bed over a short time interval becomes:

$$m = -\frac{dX}{d\tau}. \quad (14)$$

Substituting Eq. 13 into Eq. 14 and integrating the equation, the drying time can be calculated as follows:

$$\tau = -\frac{X_{cr}}{m_0} \int_{X_{R1}}^{X_{R2}} \frac{dX_R}{f(X_R)}, \quad (15)$$

where X_{cr} is the critical moisture content. In this dissertation, the constant drying rate m_0 is always given per dry mass of the bed [$\text{kg}_{\text{H}_2\text{O}}/(\text{kg}_{\text{db}} \cdot \text{s})$]. If Eq. 15 can be integrated, the drying time can be analytically calculated. If Eq. 15 cannot be integrated, the drying time must be numerically solved. When Eq. 15 is used for the fixed bed, the constant drying rate can be calculated as follows:

$$m_0 = \frac{\alpha a}{\rho_{bulk} \cdot l_v} \cdot \frac{t_{air,in} - t_{air,out}}{\ln \frac{t_{air,in} - t_{wb}}{t_{air,out} - t_{wb}}}, \quad (16)$$

where t_{wb} is the surface temperature of the material during the constant drying rate period, which is the same as the wet bulb temperature [°C], $t_{air,in}$ is the temperature of air before the bed, $t_{air,out}$ is the temperature of air after the bed, ρ_{bulk} represents the bulk density of dry material [kg_{db}/m³] and l_v is the vaporization heat of water at t_{wb} [J/kg]. In Eq. 16, the product αa represents the volumetric heat transfer coefficient. [III]

3.3.3 Economic analysis for soot sludge and sawdust mix

The economic analysis conducted in Publ. III assesses the economic feasibility of drying the soot sludge in a fixed bed belt dryer before the combustion. Drying the soot sludge is an energy efficiency investment. The economy of the energy efficiency investment is typically evaluated using the payback period without interest (Eq. 17). The characteristic drying curve was used for calculation of the drying time of the specific drying process in an economic analysis. The payback period for the dryer is

$$PBP = \frac{I_{tot,dryer}}{S_{net}}, \quad (17)$$

in which $I_{tot,dryer}$ is the total investment costs [€] (Eq. 4), and S_{net} denotes the annual net savings when the soot sludge and sawdust mixture is dried, instead of the direct combustion of moist soot sludge [€/a]. Savings result from reduced heavy fuel consumption and lower CO₂ emissions. The rate of interest is 0 % in the calculations. The operational costs of the dryer and purchased cost of sawdust result in some additional expenses. When savings and additional expenses are taken into account, net savings become

$$S_{net} = P_{savings,fuel} + P_{savings,CO_2} - A_{oper} - P_{sawdust}. \quad (18)$$

The savings, in Eq. (18), are calculated as follows:

$$P_{savings,fuel} + P_{savings,CO_2} = \Delta \dot{m}_{fuel} \cdot \gamma \cdot C_{fuel} + \Delta \dot{m}_{fuel} \cdot \gamma \cdot q_{oil} \cdot S_{CO_2} \cdot C_{CO_2}, \quad (19)$$

where $\Delta \dot{m}_{fuel}$ is the reduced heavy fuel oil consumption [kg/h], γ denotes the annual operating time of the dryer [h/a], C_{fuel} is the price of the heavy fuel oil [€/t-oil], q_{oil} is the heating value of heavy fuel oil [MJ/kg-oil], S_{CO_2} is CO₂ emission factor of heavy fuel oil [kg-CO₂/MWh] and C_{CO_2} represents the price of the CO₂ emissions [€/t-CO₂]. The reduced heavy fuel oil consumption is calculated as follows [II]:

$$\Delta \dot{m}_{fuel} = \frac{(u_1 - u_2) \cdot 2.443 + r \cdot q_{sawdust} - 2.443 \cdot r \cdot u_2}{q_{oil}} \cdot \dot{m}_{soot} \quad (20)$$

where $\dot{m}_{\text{soot}}$ is the mass flow rate of the soot sludge [kg_{db}/s], u_1 is the moisture content of the moist soot sludge [kg_{H2O}/kg_{db}], u_2 denotes the moisture content of the soot sludge and sawdust mixture after drying, r is the mass ratio (dry basis) of sawdust to soot sludge in the mixture [-] (This value is known from the drying data) and q_{sawdust} is the lower heating value of the sawdust [MJ/kg_{db}]. The term 2.443 represents the vaporization heat of water at 25 °C.

The annual operational costs of the dryer (A_{oper} , Eq. 6) include both the electricity and the heat costs. The annual cost of the sawdust is determined as follows:

$$P_{\text{sawdust}} = \dot{m}_{\text{sawdust}} \cdot \gamma \cdot C_{\text{sawdust}} \cdot q_{\text{sawdust}}, \quad (21)$$

where $\dot{m}_{\text{sawdust}}$ is the mass flow rate of the sawdust into the dryer [kg_{db}/s] and C_{sawdust} represents the price of the sawdust [€/MWh]. [III]

4. Results and discussion

4.1 Publication I: Drying of woodchips

4.1.1 Initial values

Four different drying air temperatures (30, 50, 70, 90 °C) and three different bed heights (100, 300, 500 mm) were used in the tests. For most tests, the air velocity of the drying air per cross-sectional area of the bed was 0.7 m/s. When this velocity was used, the bed remained a fixed bed without the problem of dust. In addition, tests with air velocities of 0.5 and 0.3 m/s were conducted using a bed height of 300 mm and air temperature of 70 °C in order to evaluate the influence of the air velocity on the drying time and pressure drop over the bed. The bulk density of woodchips was 143 kg_{db}/m³. The bulk density represents the average of the woodchips used in the tests (see Fig. 4). Table 8 shows the initial values for all experimental tests produced with woodchips.

The relative differences (Eq. 1) varied in the tests from 0.5 to 4.5 %. The greatest relative differences were observed in measurements with lower bed heights.

Table 8. The initial values of all experimental test conducted with woodchips. [1]

Bed height (mm)	Inlet air temperature (°C)	Absolute humidity of the inlet air (g _{H2O} /kg _{Dry Air})	Air velocity per free cross-sectional area of the drying chamber (m/s)	Initial moisture content of wood chips in the tests (kg _{H2O} /kg _{Dry Basis})	Number of tests
100	30	2	0.7	1.34; 1.34	2
100	50	2	0.7	1.27; 1.27	2
100	70	2	0.7	1.27; 1.40	2
100	90	4	0.7	1.34; 1.34	2
300	30	3-4	0.7	1.35; 1.44	2
300	50	3-4	0.7	1.40; 1.54	2
300	70	3	0.7	1.40; 1.40; 1.55	3
300	90	4	0.7	1.91; 1.91	2
500	30	4	0.7	1.50; 1.52	2
500	50	3	0.7	1.80; 1.80	2
500	70	2-3	0.7	1.40; 1.74	2
500	90	4	0.7	1.46; 1.49	2
300	70	2	0.5	1.57; 1.58	2
300	70	2	0.3	1.63; 1.63	2

4.1.2 Results: Drying kinetics and key observations

Figure 11 shows the actual drying curves (see Section 3.2.1) of woodchips for bed heights of 100, 300, and 500 mm with an air velocity of 0.7 m/s (Figs. 11A-C). Fig. 11D shows the actual drying curves of woodchips for a bed height of 300

mm and inlet air temperature of 70 °C when different air velocities (0.7, 0.5, and 0.3 m/s) have been used. For all drying curves in Fig. 11, the initial moisture content of woodchips is $\sim 1.25 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (56 % wb) and the final moisture content is $0.2 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (17 % wb). The initial moisture content of the woodchips (56 %) was selected because the drying curves (Fig. 11) could be formed for this maximum initial moisture content.

Figure 11. Drying curves of woodchips for the bed heights of 100, 300, and 500 mm (11A, 11B, and 11C, respectively), in which the air velocity has been 0.7 m/s. In 11D, drying curves (0.7, 0.5, and 0.3 m/s) of woodchips are shown for a bed height of 300 mm and inlet air temperature of 70 °C. [1]

The pressure drop over the woodchip bed was also measured in the experimental tests. Table 9 shows the pressure drop for different air velocities and bed heights. The pressure drop values in Table 9 were taken as average values from the experimental data, since the initial values (temperature and air velocity) of a certain test were stabilized. This occurred a couple of minutes after the beginning of the test. Figure 12 shows the pressure drop graphs of three tests (100, 300, and 500 mm) in which the air temperature has been 90 °C and the air velocity was 0.7 m/s.

Table 9. Experimentally measured pressure drops over the woodchip beds. [1]

Air velocity before the bed [m/s]	Bed height [mm]	Temperatures of the drying air [°C]	Pressure drop over the bed [Pa]
0.3	300	70	90–100
0.5	300	70	170–190
0.7	100	30, 50, 70, 90	70–140
0.7	300	30, 50, 70, 90	150–220
0.7	500	30, 50, 70, 90	310–470

Figure 12. The pressure drop graphs of the woodchip tests in which air temperature was 90 °C, air velocity 0.7 m/s, and bed heights were 100, 300, and 500 mm. [1]

The drying curves (Fig. 11) show that the temperature greatly influences the drying time; the lower the temperature of the drying gas, the longer the drying time needed. A similar observation has been reported in [20] and [40] where drying rates of pine bark, and sawdust and woodchips in fixed bed were studied, respectively. Figures 11A-C show that the drying time decreases considerably when the inlet air temperature increases from 50 to 70 °C. For example, for the bed height of 300 mm, drying time is almost two times longer when the inlet air temperature decreases from 70 °C (~2.2 h) to 50 °C (4.0 h). When the drying air temperature increases from 70 to 90 °C, the difference is no longer that remarkable. The bed height also affects the drying time. For example, when the bed height increases from 100 to 300 mm, the drying time with the air at 70 °C rises from ~1.3 h to 2.2 h.

Figure 11A shows that the drying rates for the bed height of 100 mm are not constantly decreasing at the end of the drying process. This most likely is a result of an uneven drying of the particles. When air flows through the thin woodchip bed and part of the particles dry, it is probable that air channels are formed inside the bed because of the movement and shrinkage of the particles. In such a case, the air does not flow evenly and moister areas are formed in the bed. When these moister areas dry, the particles again mix and the drying process

becomes more effective. It is also possible that the woodchip bed itself might contain larger particles (for example, a longer stump of branch) or smaller, easily flying particles that create the channels inside the bed. Figures 11B and C show that for the bed heights of 300 and 500 mm, the drying rates decrease more constantly through the whole drying process because formation of channels is more difficult for thicker beds. Drying curves presented in [20] and [50] also show similar behaviour even though wood species have been different in these tests, willow chips in [50] and pine bark in [20]. Results in [40] show also that drying curves for woodchips behave in the same way as they do in our tests. However, drying times are approximately 3-22 % shorter for woodchips in our studies.

Figure 11D reveals that increasing the air velocity from 0.3 to 0.5 m/s has a remarkable impact on the drying time (3.7 h versus 2.2 h). In contrast, the difference in drying times between air velocities of 0.5 and 0.7 m/s is quite small (2.2 h and 1.9 h, respectively). This indicates that it would be reasonable to use an air velocity of at least 0.5 m/s. The reason for drying times remaining at the same magnitude with air velocities of 0.5 and 0.7 m/s can be explained by the way drying progresses inside the bed. In the beginning of the drying process, material dries in the upper part of the bed with condensation simultaneously occurring in the bottom part of the bed. Condensation occurs because the air is fully saturated before it exits the dryer. As drying proceeds, condensation ends at a certain stage because less water is transferred from the material to the air in the upper part of the bed, and material is warmer in the bottom part of the bed. Yrjölä and Saastamoinen [87] have also noticed that water condensation may occur in the beginning of drying when material is dried in a fixed bed. The water absorption capacity of air through the bed is dependent on the mass flow rate of the air (inlet air temperature is the same for each velocity in Fig. 11D). When air velocities are higher (0.7, 0.5 m/s), mass flow rates also become higher. A higher water absorption capacity is the primary reason for drying times becoming lower for air velocities of 0.5 and 0.7 m/s than for the air velocity of 0.3 m/s.

According to the results, air velocity does not have such a remarkable influence on the drying time when it rises from 0.5 to 0.7 m/s. This indicates that the water (or mass) transfer rate from material to the drying air has a more significant effect on the drying time than the water absorption capacity when the air velocity exceeds 0.5 m/s. The water transfer from the material to the air depends on the mass transfer resistances inside the material and in the boundary layer. The air velocity only impacts the mass transfer resistance in the boundary layer. This resistance seems to be clearly lower than the resistance inside the material, thus explaining the reason for the drying time not becoming significantly shorter with higher air velocities.

Table 9 indicates that the pressure drops vary between the tests with the same initial values (air velocity, air temperature, and bed height). Due to the very heterogenous nature of woodchips, particle size distribution has varied between tests with the same initial values explaining the variation of pressure drops. The pressure drops in Fig. 12 show both the unstable first part of the test and the

stabilized part. It can be observed that the pressure drop constantly decreases as the drying proceeds (Fig. 12). The behaviour of the pressure drop probably results from the slight movement of the chips in the bed as they dry as well as from the increase of void volume in the bed. Studies in [87] and [40] also states that the pressure drop decreases as the drying process proceeds. However, Yrjölä and Saastamoinen [87] state that the pressure drop decreases during drying because of shrinkage of the bark and woodchip particles. Bengtsson [40] concludes in his paper that the pressure drop decreases during drying due to the decreasing air flow resistance that is a result of the reduced moisture content of wood particles. Yrjölä and Saastamoinen [87] have also mentioned that reduction of the pressure drop during drying exists probably only with the fixed beds because of the increase of the void volume in the bed. The pressure drop over the bed height of 300 mm is around 90–100 Pa with an air velocity of 0.3 m/s. When air velocities have been 0.5 and 0.7 m/s, pressure drops have fluctuated in the same magnitude of 200 Pa (Table 9). This result indicates that when the air velocity increases, channels are formed more easily inside the bed after a certain velocity; therefore, the pressure drop does not rise linearly with increasing air velocity. This means that using a higher air velocity in a dryer does not necessarily increase the electricity consumption of fans so much.

4.2 Publs II and III: Drying of soot sludge and sawdust mix

4.2.1 Initial values

Experimental tests were conducted for soot sludge and sawdust mix using three bed heights (100, 200 and 300 mm), three drying air temperatures (40, 80, 100 °C), and an air velocity of 0.75 m/s. Additionally, an air velocity of 0.9 m/s was also used in a measurement. These air velocities were chosen for the tests because higher air velocities would increase the dusting problem of the mixture when the mixture dries. Air velocity refers to the velocity before it reaches the bed. The air temperatures of 40, 80 and 100 °C were chosen for the measurements because temperatures of waste heat streams are usually high enough to heat the drying air to these temperatures. The optimal mixture ratio for the mixture was first studied. The mixture ratios (vol.-%) for the soot sludge and sawdust used in the first tests were 30%_{soot}:70%_{sawdust}, 50%_{soot}:50%_{sawdust} and 70%_{soot}:30%_{sawdust}. A summary of all the tests conducted for soot sludge and sawdust mixture is shown in Table 10. The moisture content values for sawdust and soot sludge shown in Table 10 were determined using the method described on page 31.

Absolute humidity of the inlet air varied between 2.5 and 5.7 g_{H₂O}/kg_{da} (see Table 10). The relative differences (Eq. 1) varied in the tests from 0.5 to 2.5 %.

The bulk densities were measured by mixing samples for each volume ratios, and measuring their bulk volumes before the oven drying. After the drying, the dry masses of the samples were weighed. The bulk densities used in [II] were calculated using these two values (dry mass/bulk-volume). The following bulk

densities were defined: 133 kg_{db}/bulk-m³ (30%_{soot}:70%_{sawdust}), 116 kg_{db}/bulk-m³ (50%_{soot}:50%_{sawdust}) and 79 kg_{db}/bulk-m³ (70%_{soot}:30%_{sawdust}).

The desired final moisture content for the soot sludge and sawdust mixture was 0.7 kg_{H₂O}/kg_{db} (40 % wb) in the calculations. This makes it possible to handle and combust it properly in a boiler without major dusting problems (information gained from the CHP plant situated in Eastman's production plant area).

Table 10. Summary of all tests conducted for the soot sludge and sawdust mixture. [II]

Z (mm)	T (°C)	Mixture ratio (vol.%)	Number of tests	U _{1, soot sludge} (kgH ₂ O/kgd.b.)	U _{1, sawdust} (kgH ₂ O/kgd.b.)	U _{1, mixture} (kgH ₂ O/kgd.b.)	V (m/s)	X (gH ₂ O/kgd.a.)
200	80	30% _{soot} , 70% _{sawdust}	2	19.62; 18.62	1.20; 1.22	3.56; 3.30	0.75	3.6; 2.5
200	80	50% _{soot} , 50% _{sawdust}	3	21.10; 21.10; 20.52	1.36; 1.36; 1.39	5.73; 5.83 5.51	0.75	2.6; 2.5; 2.6
200	80	70% _{soot} , 30% _{sawdust}	2	21.10; 21.10	1.36; 1.36	9.12; 9.11	0.75	2.5; 3.0
200	40	50% _{soot} , 50% _{sawdust}	1	20.40	1.26	5.26	0.75	5.7
200	100	50% _{soot} , 50% _{sawdust}	3	20.52; 21.49; 21.75	1.39; 1.52; 1.17	5.51; 6.23; 5.28	0.75	5.4; 5.5; 5.5
100	80	50% _{soot} , 50% _{sawdust}	2	20.40; 20.40	1.26; 1.26	5.30; 5.39	0.75	3.0; 2.8
300	80	50% _{soot} , 50% _{sawdust}	1	20.50	1.24	5.21	0.75	3.1
200	80	50% _{soot} , 50% _{sawdust}	1	20.52	1.39	5.75	0.9	4.1
300	40	50% _{soot} , 50% _{sawdust}	1	20.50	1.24	5.15	0.75	4.6
300	100	50% _{soot} , 50% _{sawdust}	1	20.50	1.24	5.14	0.75	5.5

4.2.2 Results: Drying kinetics and key observations

Drying curves

After defining the optimal mixture ratio of soot sludge and sawdust and the air velocity, these values were used to study the influence of the bed height and inlet air temperature on the drying curve in [II]. The optimal mixture ratio was selected in such a way that as much as soot sludge as possible could be dried with a suitable drying time. Fig. 13 shows the drying curves for different soot sludge and sawdust mixtures when the mixture ratios (vol.-%) are 30%_{soot}: 70%_{sawdust}, 50%_{soot}: 50%_{sawdust}, and 70%_{soot}: 30%_{sawdust}.

Figure 13. Drying curves for mixture ratios (vol.-%) of 30%_{soot}: 70%_{sawdust}, 50%_{soot}: 50%_{sawdust}, 70%_{soot}: 30%_{sawdust}. Bed height is 200 mm, air velocity 0.75 m/s and inlet air temperature 80 °C. [II]

Fig. 13 shows that a larger amount of soot sludge notably increases drying time to the final moisture content. It can be seen that the final moisture content of 1.3 kg_{H2O}/kg_{db} (57 % wb) is reached in around 7.4 h when the mixture ratio of 70%_{soot}: 30%_{sawdust} is used. This is an extremely long drying time for a continuous belt dryer, and the desired final moisture content of 0.7 kg_{H2O}/kg_{db} is not even achieved. Nevertheless, with the volume ratios of 30%_{soot}: 70%_{sawdust} and 50%_{soot}: 50%_{sawdust}, the desired final moisture content can be achieved, and drying times (approx. 1.5 and 2.3 h, respectively) are realistic for a continuous belt dryer. As these drying times are of the same magnitude, it is reasonable to choose the volume ratio of 50%_{soot}: 50%_{sawdust}, in which case, more soot sludge can be dried at once.

It was noted in the tests that the material did not dry evenly when the volume fraction of soot sludge was 70 %. This is probably the main reason for the lengthier drying time for the mixture of 70%_{soot}: 30%_{sawdust} compared to other ratios. When the ratio was 70%_{soot}: 30%_{sawdust}, it was observed that the bed material had shrunk and considerably cracked on the top of the bed during drying, but the bottom part was still moist. The bed was still like an unbroken

cake, even though it was cracked on the top (see Figs. 14, 15a and b). This phenomenon did not occur with the other volume ratios.

Figure 14. A shrunken and cracked material cake inside the dryer after drying when the mixture ratio was 70%_{soot}: 30%_{sawdust}. [II]

Figure 15. Moist and dried soot sludge and sawdust mixture: a) moist mix with mixture ratio of 70%_{soot}: 30%_{sawdust}, b) dried mix with mixture ratio of 70%_{soot}: 30%_{sawdust}, c) moist mix with mixture ratio of 50%_{soot}: 50%_{sawdust}, d) dried mix with mixture ratio of 50%_{soot}: 50%_{sawdust}. [II]

When drying 30%_{soot}: 70%_{sawdust} and 50%_{soot}: 50%_{sawdust} beds, the cracking of the bed occurred considerably more, forming small, semi-solid, and dry pieces of the mixture (see Fig. 15d). Fig. 15c and d show moist (before drying) and dry (after drying) mixtures for the mixture ratio of 50%_{soot}: 50%_{sawdust}, respectively. Although the bed contains more water when the mixture ratio is 70%_{soot}: 30%_{sawdust}, the caking effect is probably the main reason for the much lengthier drying time for this mixture ratio compared to the other ratios. The drying air cannot flow evenly through the extremely moist bed; thus, wetter and denser areas are formed, which notably increase the drying time.

Figure 15a shows that the mix has a very solid nature before drying as a result of high water content. In practice, it is impossible to notice any sawdust particles in the mix. The mixture was also very sticky when it was loaded into the dryer. Fig. 15b shows that the mixture still has a solid nature after drying due to the above-mentioned caking effect. However, sawdust particles can clearly be observed in the mix. With a mixture ratio of 70%_{soot}: 30%_{sawdust}, the mixture was not overly dusty and sticky after drying. However, similar kind of crust phenomenon, which was noted in [30] for pure activated sludge, was detected for the mixture of 70%_{soot}: 30%_{sawdust} after drying. With a mixture ratio of 50%_{soot}: 50%_{sawdust}, the soot sludge and sawdust mix more closely resembles a mixture even before drying (Fig. 15c). After drying, the consistency of the mix did not change much compared to the situation before drying, but the mix no longer retained its sticky nature. Instead, it is a rather dusty material. It was also possible to see that some shrinkage in the bed had occurred during drying. However, it was beyond the scope of the study to explicitly define the shrinkage of the bed. For example, Li et al. studied shrinkage of the bed in the convective drying of sewage sludge and sawdust in [27]. Bennamoun et al. [25] also identified the same kind of shrinkage and cracking effect as mentioned before in municipal sludge drying. Li et al. [88] noticed a similar shrinkage effect in drying of a sewage sludge and sawdust mixture.

The results of this dissertation of soot sludge and sawdust mixture drying supports results noted in [26, 27]: mixing a specific amount of sawdust to a sludge can enhance the drying process/ drying rate. However, the difference between the mass ratios (expressed on a dry matter basis for both the sludge and sawdust) is significant. In [26, 27], sawdust addition is shown to have a positive impact on the drying process from a mass ratio of 2/8 (sawdust/wastewater sludge). For sawdust and soot sludge mixture used in this dissertation, the most suitable mass ratio (db) is 7.7/2.3 (50%_{soot}: 50%_{sawdust}).

The drying curves for air velocities of 0.75 and 0.9 m/s are shown in Fig. 16a. Increasing the air velocity slightly improves the drying rate. As expected, drying with an air velocity of 0.9 m/s is faster, but the difference between drying times is less than half an hour when the final moisture content is $\sim 0.3 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$. A slightly shorter drying time when drying with an air velocity of 0.9 m/s mainly results from the greater mass flow rate through the bed. Myllymaa et al. have also found these kinds of results for woodchip drying in a fixed bed [49]. Results indicate that it is not worthwhile using a higher air velocity than 0.75 m/s,

because drying times for the desired final moisture content of $0.7 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ are at the same magnitude. A higher air velocity also increases the electricity consumption of the fans.

The influence of bed height on drying time was studied using three different bed heights when the mixture ratio was $50\%_{\text{soot}}: 50\%_{\text{sawdust}}$. Fig. 16b presents drying curves for these bed heights (100, 200 and 300 mm). The corresponding cross-sectional areas (Eq. 5) for these bed heights become 26 m^2 (100 mm), 18 m^2 (200 mm), and 20 m^2 (300 mm), when the final moisture content is $\sim 0.7 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ and the mass flow of the moist soot sludge is 20 t/day . On the basis of this result, the most optimal bed height is 200 mm, because the cross-sectional area is smaller than for other bed heights. The pressure drop over the bed is also lower for a bed height of 200 mm than for a bed height of 300 mm, which reduces operational costs. The pressure drops over the bed are discussed in more detail at the end of this section.

The tests also showed that the material dried quite evenly, despite the different bed height. The material after drying with a mixture ratio of $50\%_{\text{soot}}: 50\%_{\text{sawdust}}$ was visually similar to the one in Fig. 15d, regardless of bed height.

Figure 16. a) Drying curves for air velocities of 0.75 and 0.9 m/s. Bed height is 200 mm, air temperature 80°C and mixture ratio $50\%_{\text{soot}}: 50\%_{\text{sawdust}}$. b) Drying curves for bed heights of 100, 200 and 300 mm. Mixture ratio is $50\%_{\text{soot}}: 50\%_{\text{sawdust}}$, air velocity 0.75 m/s and temperature 80°C . [11]

The influence of the inlet air temperature on drying time was determined for bed heights of 300 and 200 mm. The drying curves of the mix for these bed heights are shown in Fig. 17. For the recommended bed height of 200 mm, cross-sectional areas for each inlet air temperatures become 41 m^2 (40°C), 18 m^2 (80°C), and 15 m^2 (100°C), when the final moisture content is $0.7 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$, and the total mass flow rate $184 \text{ kg}_{\text{db}}/\text{h}$ (moist soot sludge: 20 t/day = mass flow of dry sludge $42 \text{ kg}_{\text{db}}/\text{h}$ and mass flow of sawdust $142 \text{ kg}_{\text{db}}/\text{h}$). For a bed height of 300 mm, the cross-sectional areas become 42 m^2 (40°C), 20 m^2 (80°C), and 16 m^2 (100°C) when the final moisture content and the mass flow rate are the same as in the case of 200 mm. These results also suggest that there is no reason to use a bed height higher than 200 mm in the dryer.

Figure 17. a) Drying curves for air temperatures of 40, 80, and 100 °C. Mixture ratio is 50 %_{soot}: 50 %_{sawdust}, air velocity 0.75 m/s and bed height 300 mm. b) Drying curves for air temperatures of 40, 80, and 100 °C. Mixture ratio is 50%_{soot}: 50%_{sawdust}, air velocity 0.75 m/s, and bed height 200 mm. [11]

Pressure drop over the bed

The pressure drop over the bed was determined in each test. Fig. 18 shows an example of the pressure drop over the bed when the bed height is 200 mm, mixture ratio 50%_{soot}: 50%_{sawdust}, air velocity 0.75 m/s and air temperature 80 °C. Fig. 18 presents the pressure drop data over the whole test, including the short initial part where the desired process values (e.g. air velocity) had not yet been stabilized. When the desired process values were stabilized, the pressure drop over the bed initially increases, because the mixture bed is tighter at the beginning of the process. As the drying progresses, air channels simultaneously form inside the bed and moist material dries and shrinks. As a result, the bed becomes looser and the pressure drop decreases. Similar results have been reported in [49, 40, 87]. For all tests, the pressure drop was typically quite high at the beginning of the drying period, and reduced rapidly as the drying progressed.

Variations of pressure drops over the bed for various initial values are shown in Table 11. The first pressure drop value in Table 11 is the maximum pressure drop of the test at the beginning, when the initial values (air velocity and temperature) stabilized, and the later value is the minimum pressure drop at the end of the test. The final moisture content was approximately 0.3–0.4 kg_{H₂O}/kg_{db} at the end of all tests, except for the mixture ratio of 70%_{soot}: 30%_{sawdust}, where the final moisture content was ~1.3 kg_{H₂O}/kg_{db}. As expected,

the pressure drop increases when the bed height becomes higher. The results in Table 11 also show that when the volume of the soot sludge in the mixture increases, the pressure drop at the beginning increases notably. Increasing air velocity does not seem to have any remarkable effect on pressure drop.

Figure 18. Pressure drop over the bed for a bed height of 200 mm. Mixture ratio is 50%_{soot}:50%_{sawdust}, air velocity 0.75 m/s, and air temperature 80 °C. [II]

Table 11. Variations of pressure drops over the bed (approximations from the measurement data). [II]

Air velocity before the bed (m/s)	Mixture ratio sludge: sawdust (vol.%)	Bed height (mm)	Temperature of the drying air (°C)	Pressure drop over the bed: at the beginning–at the end of tests (Pa)
0.75	50: 50	300	40, 80, 100	585–130
0.75	50: 50	200	40, 80, 100	550–85
0.75	50: 50	100	80	250–80
0.9	50: 50	200	80	530–190
0.75	70: 30	200	80	830–65
0.75	30: 70	200	80	510–210

Characteristic drying curves

Characteristic drying curves were presented in Publ. III using the equations in Section 3.3.2. Figures 19 a and b show characteristic drying curves for soot sludge and sawdust mix for the bed heights of 200 and 300 mm, and air temperatures of 40, 80 and 100 °C. Figure 19 reveals that the shape of the characteristic drying curve is almost linear for both bed heights, regardless of the temperature level. This indicates that the relative drying rate is independent of the level of the inlet air temperature over the temperature range of 40 – 100 °C. Furthermore, results indicate that characteristic drying curves for the bed heights of 200 mm and 300 mm could also be used to define the drying time when the air temperature is to some extent over 100 °C. Fig. 19c presents an almost linear characteristic drying curve for the air velocities of 0.75 and 0.9 m/s. It is reasonable to conclude that the relative drying rates are quite independent of the air velocity over the velocity range used in this study for the bed height of 200 mm. The shape of the characteristic drying curve of the soot

sludge and sawdust mixture resembles those found in previous studies for several materials [21, 89-92].

Figure 19. a and b) Characteristic drying curves for the bed heights of 200 mm and 300 mm. Air velocity is 0.75 m/s in all cases. Mixture ratio is 50%_{soot}: 50%_{sawdust}. c) Characteristic drying curve for the air velocities of 0.75 and 0.9 m/s. Bed height is 200 mm and air temperature 80 °C in both cases. Mixture ratio is 50%_{soot}: 50%_{sawdust}. [III]

The accuracy of the characteristic drying curves (Eqs. 22, 23 and 24) was evaluated by comparing the drying times calculated using the characteristic drying curves (Eq. 25), and experimentally measured drying times from earlier studies conducted by the authors [II]. In the drying time calculations, the constant drying period was included in the drying time as Eq. 25 shows. The critical moisture contents (X_{cr}) and the constant drying rates (m_0) were defined for every bed height and air temperature combination from the original drying curves.

$$f(x_R) = 0.9559x_R + 0.0495 \quad (22)$$

$$f(x_R) = 0.9958x_R - 0.0064 \quad (23)$$

$$f(x_R) = 0.9804x_R + 0.0248 \quad (24)$$

$$\tau = -\frac{X_{cr}}{m_0} \int_{X_{R1}}^{X_{R2}} \frac{dx_R}{f(x_R)} = -\frac{X_{cr}}{m_0} \cdot \left[\int_{X_{R1}}^1 dx_R + \int_1^{X_{R2}} \frac{dx_R}{f(x_R)} \right] \quad (25)$$

Table 12 shows the results of the comparison. The relative errors between the drying times vary from 0.5 and 2.7 % when the characteristic drying curves have been used for the bed heights of 200 mm and 300 mm. For air velocities of 0.75

and 0.9 m/s, the relative errors between the drying times range from 2.6 to 3.8 %. In general, these results signify that the characteristic drying curves (Eqs. 22 – 24) estimate drying times with a good accuracy. Table 12 also shows the critical moisture contents of the specific drying processes used in CDC calculations. The values for the term m_0 can be seen in Tables 13 and 14.

Table 12. The comparison of the experimentally measured drying times (DC) and the calculated drying times using the characteristic drying curves (CDC). Initial and final moisture contents have been 5.3 (84 %) and 0.7 kg_{H2O}/kg_{db} (40 %) in all cases. The relative errors have been defined as the following: $|1 - \text{Time}_{\text{CDC}} / \text{Time}_{\text{DC}}| * 100 \%$. [III]

The comparison of the drying times						
	200 mm			300 mm		
	Drying time, DC [s]	Drying time, CDC [s]	Relative error	Drying time, DC [s]	Drying time, CDC [s]	Relative error
40 °C	17870	18347	2.7	27930	28308	1.4
80 °C	7840	7790	0.6	13010	13219	1.6
100 °C	6650	6526	1.9	10360	10408	0.5
0.75 m/s; 80 °C	7840	8137	3.8			
0.9 m/s; 80 °C	7650	7453	2.6			
The critical moisture contents X_{cr} [kg _{H2O} /kg _{db}]						
	200 mm; 0.75 m/s			300 mm; 0.75 m/s		
40 °C	4.50			4.16		
80 °C	5.04			4.34		
100 °C	4.30			4.46		
	200 mm; 0.9 m/s					
80 °C	4.51					

Volumetric heat transfer coefficient for the CDC

A value for the term αa was calculated from the experimental data in Publ. III. Measuring the outlet air temperature during the constant drying rate period, and assuming that the surface temperature of the particles in the bed was the wet bulb temperature, the term αa could be solved from Eq. 16. Tables 13 and 14 show the volumetric heat transfer coefficients for the bed heights of 200 and 300 mm, respectively. The air temperatures before the bed ($t_{\text{air, in}}$) and after the mixture bed ($t_{\text{air, out}}$) were experimentally measured, and the values were taken as average values when the drying was in the period of the constant drying rate.

The average values of the term αa can be used to calculate the constant drying rate (m_0 , Eq. 16) for the soot sludge and sawdust mix in a fixed bed when the inlet air temperature is ≤ 100 °C. The average αa value can also be used to estimate the constant drying rate although the inlet temperature is slightly higher than 100 °C if more accurate data is not available.

Results show that the volumetric heat transfer coefficient becomes lower when the bed height increases. The obvious reason for this is that vapour from the drying air condenses in the bottom part of the bed (air moves from top to down); therefore, the drying rate (m_0) remains lower for higher bed heights. In fact, there is no constant drying rate period over the whole bed when the bed height is 300 mm. In general, this is quite normal behaviour in the fixed beds when the bed heights are sufficiently high and mass transfer resistances from material to air are low.

Table 13. The calculated values for the term αa for the bed height of 200 mm. Bulk density of mixture is 116 kg_{db}/m³ in all cases. [III]

Air velocity: v [m/s]	$t_{\text{air,in}}$ [°C]	$t_{\text{air,out}}$ [°C]	x_{in} [kg _{H2O} /kg _{da}]	t_{wb} [°C]	m_0 [kg _{H2O} /(kg _{db} s)]	αa [W/(m ³ K)]
0.75 m/s	40.9	21.3	0.0057	19.4	0.000461	16160
0.75 m/s	81.4	31.1	0.0026	27.6	0.001190	18274
0.75 m/s	101.7	37.4	0.0055	32.9	0.001249	14895
Average:						16443
0.9 m/s	81.5	34.5	0.0041	28.4	0.001177	15309

Table 14. The calculated values for the term αa for the bed height of 300 mm. Bulk density of mixture is 116 kg_{db}/m³ and air velocity is 0.75 m/s in all cases. [III]

$t_{\text{air,in}}$ [°C]	$t_{\text{air,out}}$ [°C]	x_{in} [kg _{H2O} /kg _{da}]	t_{wb} [°C]	m_0 [kg _{H2O} /(kg _{db} s)]	αa [W/(m ³ K)]
40.7	20.0	0.0046	18.5	0.000308	11435
80.0	31.3	0.0031	27.6	0.000685	10534
103.4	35.1	0.0055	33.2	0.000893	13266
Average:					11745

4.2.3 Results: Economic analysis for sludge and sawdust mix

The economic analysis was carried out for the soot sludge and sawdust mix [III] using the equations presented in Section 3.3.3.

Combustion of the soot sludge requires heavy fuel oil as a supporting fuel in a boiler. Drying converts the negative LHV of the sludge in a positive one. It was assumed in the calculations that the moist soot sludge is dried with sawdust in a continuous belt dryer, and the dried mixture is combusted in a boiler. It was also assumed that the fuel input into the boiler remains the same. As a result of drying, heavy fuel consumption reduces in the boiler.

The analysis was only conducted for the bed height of 200 mm. Inlet air temperatures in the analysis were 40, 60, 80, 100 and 120 °C.

Payback periods were first calculated for the base case study. In the base case study, the air temperatures were 40, 60, 80, 100 and 120 °C. For the air temperatures of 60 and 120 °C, the characteristic drying curve was used to define the drying times (Eq. 22 and Eq. 25). For other temperatures, the drying

times were defined using the experimental data. When the characteristic drying curve was used, the critical moisture content was $4.62 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$, and the outlet air temperature (Eq. 16, $t_{\text{air,out}}$) was calculated as follows: $t_{\text{air,out}} = t_{\text{wb}} + 3.3^\circ\text{C}$. Both values were extracted from the experimental data [II]. Table 15 summarizes all the initial values for this economic analysis.

Table 15. The initial values for the economic analysis. [III]

Parameters	Initial value
Volume ratio of the sludge and sawdust mixture	50% _{soot sludge} : 50% _{sawdust}
Bed height of the mixture inside the dryer	200 mm
Inlet air velocity before the bed	0.75 m/s
Mass flow rate of moist soot sludge	20 t/d
Moisture content of pure soot sludge	95 % wb
Mass flow rate of moist sawdust	7.8 t/d
Moisture content of pure sawdust	56 % wb
Initial moisture content of the mixture	$5.3 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (84 % wb)
Final moisture content of the mixture after drying	$0.7 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (40 % wb)
Bulk density of the mixture	$116 \text{ kg}_{\text{db}}/\text{m}^3$
Total pressure drop of the dryer for the bed height of 200 mm	318 Pa (from the drying data, average value between the pressure drops in the beginning and at the end of the tests [II])
Total pressure drop over the heat exchanger and air duct	400 Pa
Drying air temperature before the fans and before heating it to the desired drying air temperature	15 °C
Moisture content of air	$0.004 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{da}}$ (average value from the drying tests [II])
Efficiency of the fan (η)	0.8
Heating value of heavy fuel oil	41 MJ/kg
Emission factor of heavy fuel oil	284 $\text{kg}_{\text{CO}_2}/\text{MWh}$
Price of the CO_2 ton	13 €/t _{CO2} (current price, [93] base case)
Lower heating value of dry sawdust	19 MJ/kg _{db}
Lower heating value of dry soot sludge	22.65 MJ/kg _{db}
Mass ratio of sawdust to soot sludge in the mixture (r)	$3.4 \text{ kg}_{\text{db, sawdust}}/\text{kg}_{\text{db, soot sludge}}$
Price of electricity	40 €/MWh
Price of heat	5 €/MWh (base case)
Price of sawdust	19 €/MWh (current price in Finland [94], base case)
Operating time of the chemical plant and the dryer	8000 h/a

Fig. 20 shows the base case results. Table 16 shows the payback period calculations for the base case when the current import price of the heavy fuel oil (410 €/t-oil, [95]) has been used. Usually, energy efficiency investments are seen as cost-effective if the payback period is 2-3 years or less (a rule of thumb based on conversations with people working in Finnish energy companies). Fig.

20 shows that drying the soot sludge and sawdust mix is profitable (payback period is 3 years or less) when the inlet air temperature is 80 °C or more, and the heavy fuel oil price is at least 250 €/t-oil. Drying is also profitable for air temperatures of 40 and 60 °C if the price of heavy fuel oil is at least 350 €/t-oil. The current import price of heavy fuel oil is 410 €/t-oil, which means that drying is economic for all inlet air temperatures. Although this analysis assumes that drying reduces oil consumption, these results are parallel/ valid for any fuel that can be combusted alongside the soot sludge. Table 16 shows that the payback period ranges from 1.0 to 2.4 a, when the current price of the heavy fuel oil is used in the calculations. Fig. 20 also confirms that there is no reason to use a higher drying air temperature than 80 °C, because the payback periods of the air temperatures from 80 to 120 °C are at the same magnitude.

Figure 20. The payback periods for the soot sludge and sawdust mixture dryers when the inlet air temperatures are 40, 60, 80, 100 and 120 °C. All initial values are shown in Table 15. [III]

Table 16. The payback period calculations for the base case when the current import price of the heavy fuel oil (410 €/t-oil, [95]) has been used. [III]

	40 °C	60 °C	80 °C	100 °C	120 °C
Cross-sectional area of the dryer	39 m ²	24 m ²	17 m ²	15 m ²	13 m ²
Investment cost	535131 €	359705 €	278879 €	244024 €	219992 €
Savings in heavy fuel oil costs	346256 €/a	346256 €/a	346256 €/a	346256 €/a	346256 €/a
Savings in CO ₂ emissions	35510 €/a	35510 €/a	35510 €/a	35510 €/a	35510 €/a
Cost of sawdust	114391 €/a	114391 €/a	114391 €/a	114391 €/a	114391 €/a
Cost of electricity	8492 €/a	5117 €/a	3726 €/a	3160 €/a	2783 €/a
Cost of heat	33355 €/a	34007 €/a	33738 €/a	35417 €/a	36571 €/a
Payback period	2.4 a	1.6 a	1.2 a	1.1 a	1.0 a

In Publication III, sensitivity analyses were also carried out for soot sludge and sawdust drying where the influence of the heat, sawdust and emission

prices on the payback period were studied. These analyses revealed that the heat, sawdust and emission prices have no remarkable influence on the economy of drying if the heavy fuel oil price does not fall below c. 300 €/t-oil. The payback period is less than three years for most cases in the sensitivity analyses.

On the basis of the economic analysis, it seems to be economical to dry soot sludge by mixing it with sawdust.

4.3 Publ IV: Techno-economic evaluation of drying in moving beds: The effect of drying kinetics on drying costs

4.3.1 Initial values

In Section 4.3, calculation results for base cases (woodchips, bark, and mixture of soot sludge and sawdust) are first presented, and then the results of sensitivity analysis shown. The equations discussed in Section 3.3.1 were used in the techno-economic evaluation. Table 17 shows the initial values for the techno-economic base case calculations. The parameters used when the dryer size and annual costs were simulated were: initial and final moisture contents, air temperature, bed height and air velocity. Table 18 summarises the parameters used in the calculations.

Table 17. Summary of the initial values for all base case calculations. [IV]

Annuity calculations		
Parameter	Term	Value
Economic lifetime of the dryer	n	20 a
Rate of interest	i	5%
Operational cost calculations		
Parameter	Term	Value
Operating time of continuous-working dryer	γ	8000 h/a
Total pressure drop over the heat exchanger and air duct	$\Delta P_{\text{he+ad}}$	400 Pa
Efficiency of the fan(s)	η_{fan}	0.8
Moisture content of air (average value from all drying tests)	x	0.004 kg _{H₂O} /kg _{da}
Temperature of drying air before the heat exchanger	t_0	15 °C
Price of heat	C_{heat}	5 €/MWh
Price of electricity	C_{elec}	40 €/MWh

Table 18. Summary of the initial values of each material: Woodchips, bark and mixture of soot sludge and sawdust. [IV] ($\rho_{bulk} = \rho_{db}$)

Parameter	Term	Woodchips	Bark	Soot sludge + sawdust mix
Drying air temperature	t	Value: 30°C 50°C 70°C 90°C	Value: 50°C 70°C 90°C 110°C	Value: 40°C 80°C 100°C
Bed height	z	100 mm 300 mm 500 mm	50 mm 150 mm 250 mm	200 mm 300 mm
Velocity of air	v	0.7 m s ⁻¹ 0.5 m s ⁻¹ 0.3 m s ⁻¹	0.75 m s ⁻¹	0.75 m s ⁻¹ 0.9 m s ⁻¹
Highest initial moisture content	u_1	1.3 kg _{H2O} /kg _{db}	1.3 kg _{H2O} /kg _{db}	5.3 kg _{H2O} /kg _{db}
Lowest final moisture content	u_2	0.2 kg _{H2O} /kg _{db}	0.2 kg _{H2O} /kg _{db}	0.7 kg _{H2O} /kg _{db}
Pressure drop over the bed	$\Delta P_{material}$	Air velocity 0.7 m s ⁻¹ ; 100 mm: 105 Pa 300 mm: 185 Pa 500 mm: 390 Pa Air velocity 0.5 m s ⁻¹ ; 300 mm: 180 Pa Air velocity 0.3 m s ⁻¹ ; 300 mm: 95 Pa	Air velocity 0.75 m s ⁻¹ ; 50 mm: 53 Pa 150 mm: 125 Pa 250 mm: 165 Pa	Air velocity 0.75 m s ⁻¹ ; 200 mm: 318 Pa 300 mm: 358 Pa Air velocity 0.9 m s ⁻¹ ; 200 mm: 360 Pa
Bulk density of dry material	ρ_{db}	143 kg _{db} /m ³	133 kg _{db} /m ³	116 kg _{db} /m ³

Inlet air temperatures and velocities were assumed to be the same for each block (see Fig. 3) when base cases and sensitivity analysis were calculated. The possibility of using blocks with different air temperatures or velocities was also evaluated from the perspective of drying costs (see Section 4.3.2).

The calculations were performed when the initial moisture contents were 1.3 (57 % wb) or 1.0 kg_{H2O}/kg_{db} (50 %) for woodchips and bark. The initial moisture content of the soot sludge and sawdust mix was 5.3 kg_{H2O}/kg_{db} (84 %) in all

calculations. The mixture ratio of soot sludge and sawdust mix was 50 vol.-% soot sludge and 50 vol.-% sawdust in all calculations. More detailed initial values are shown in IV.

The final moisture contents for woodchips and bark were $0.2 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (17 %) and $0.6 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (38 %) in the calculations. The final moisture content for the mix of sludge and sawdust was $0.7 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (41 %). Selecting the final moisture contents has been explained in more detail in IV. The final moisture contents of material used in the calculations are the average moisture content of the bed. In reality, there is a distribution of the moisture content inside the bed. However, it was beyond the scope of this study to determine the moisture content distribution inside the material bed.

The mass flow rate of bark and woodchips was $2.64 \text{ t}_{\text{db}}/\text{h}$ in all calculations. When this value was used, the dryer size never exceeded the highest available dryer size (480 m^2) (see Section 3.3.1). In reality, the mass flow rate of material can be higher, thus indicating that the final moisture content has to be left at a higher level or, if low moisture content is desired, several dryers are needed. Naturally, multiple dryers mean higher investment costs and could affect drying cost results. The mass flow rate of sludge and sawdust mix was $0.18 \text{ t}_{\text{db}}/\text{h}$ in the calculations. For more information, see IV.

The residence times (τ , Eq. 5) of waste materials in the dryer were based on experimentally measured drying curves/data published earlier in [I, II, 21]. Drying curves of woodchips used in Publ. IV are shown in Fig. 11 of this dissertation. Respectively, drying curves of bark can be found in [21, page 956: Fig. 3 and 4, page 957: Fig. 5 and 6]. Drying curves of soot sludge and sawdust mix are shown in Figs. 16a and 17 of this dissertation. All residence times used in the calculations of IV are summarised in Attachment 1 of IV. The absolute humidity of the inlet air fluctuated between 2 and $5.7 \text{ g}_{\text{H}_2\text{O}}/\text{kg}_{\text{da}}$ in the experimental tests [I, II, 21]. This air humidity has generally been equivalent to the air humidity of the outdoor air.

The belt dryer was a continuous one in all cases. It was also assumed that there is always a sufficient amount of excess heat available in all calculation cases.

4.3.2 Results and key observations

All the results presented in this Section 4.3.2 were formed using the following method. For each drying air temperature and bed height (or air velocity) combination, the total adjusted annual drying costs were calculated using the annuity method (Eq. 2). The results (Section 4.3.2) are presented as relative drying costs by dividing the drying costs of each combination by the highest drying costs of the studied case. Number 1 in each figure always represents the most expensive dryer, and the lowest value the most economic option. However, it should be noted that the equation for the investment costs of a dryer (4) provides only an estimate of actual drying costs.

This section first presents the relative drying costs for each waste stream when air temperature, bed height, and the change of moisture content were changed (base cases). After this, the economic sensitivity analyses for rate of interest and

the prices of heat and electricity are presented. Block studies are also presented at the end of this section.

4.3.2.1 Sensitivity analysis for base cases—Initial and final moisture content

Woodchips

Figure 21 shows the effect of the bed height, drying air temperature and change of moisture content for the relative drying costs of woodchips.

Figure 21. The relative drying costs for woodchips. Air velocity is 0.7 m/s in all cases. [IV]

Figures 21 a–c reveal that increasing the bed height practically reduces drying costs for all temperatures, despite the change of moisture content. However, when the change of moisture content is 1.3–0.6 kg_{H₂O}/kg_{db} and the air temperature 30 °C (Fig. 21c), the bed height of 300 mm yields slightly lower drying costs than the bed height of 500 mm, but the difference is small. Usually a higher temperature results in lower drying costs [45]. This is an expected result because a higher air temperature reduces the dryer size and the investment costs. Nonetheless, there are a few exceptions to this trend in Fig. 21. When the air temperature rises from 30 to 50 °C, drying costs increase in Figs. 21a and b. Attachment 1 of IV shows the exact investment and operational costs for all cases shown in Figure 21. As Attch. 1 reveals, investment costs are smaller for the air temperature of 50 °C, but operational costs become higher for 50 °C, thus explaining this behaviour. A higher air temperature results in a lower mass flow rate of air, but the temperature or the enthalpy change in the heating unit increases, respectively (see Eqs. 8 and 9). For the air temperature of 50 °C, the mass flow rate does not sufficiently decrease for it to compensate for the enthalpy increase in the heating unit; therefore, the operational cost becomes higher compared to the air temperature of 30 °C. If the air temperature still increases from 50 to 90 °C, drying costs are reduced with all bed heights as expected. Figure 21 also shows that the differences between drying costs can be significant depending on the bed height and air temperature. This indicates that it is extremely important to select the correct design parameters for a belt dryer.

Figure 22 shows the effect of the initial and final moisture contents of woodchips on drying costs when different air velocities were used.

Figure 22. The relative drying costs for woodchips when different air velocities are used. Bed height is 300 mm in all cases. [IV]

The lowest drying costs are always achieved when the air velocity is 0.5 m/s. The highest drying costs vary between the air velocities of 0.3 and 0.7 m/s depending on the moisture content change. When woodchips are dried to the lower moisture content (0.2 kg_{H₂O}/kg_{db}, see Figs. 22a and b), drying costs are highest with the highest air velocity. This indicates that when the final moisture content is low, moisture removal from material is mostly restricted by internal mass transfer resistance. Higher air velocity improves external mass transfer and water absorption capacity of air, but these factors have a negligible influence on moisture removal at low moisture contents due to high internal mass transfer resistance.

However, when the final moisture content is higher (see Fig. 22c), external mass transfer resistance and the water absorption capacity of air have a greater impact on the drying rate over the bed, thereby further lowering drying costs for the air velocity of 0.7 m/s, than for the air velocity of 0.3 m/s.

In conclusion, it is not reasonable to use high air velocities in a belt dryer if material with high internal mass transfer resistance needs to be dried. This usually means material with a larger particle size, such as woodchips (see Fig. 4). If different air velocities are used in dryer blocks (see Fig. 3), air velocities should be lower in the blocks where the moisture content of the material has already been reduced.

Bark

Figure 23 presents the effect of the bed height, drying air temperature and change of moisture content for the relative drying costs of bark. Experimental tests were not conducted for several bed heights for initial moisture content of 1.3 kg_{H₂O}/kg_{db} in [21]; hence, Fig. 23 shows results only for the bed height of 250 mm.

Figure 23. The relative drying costs for bark. Air velocity is 0.75 m/s in all cases. [IV]

Figures 23a–c mainly show that by increasing the air temperature from 50 to 110 °C, the lowest drying costs are achieved with the highest bed height. However, increasing the temperature from 90 to 110 °C has no noticeable impact on drying costs.

Furthermore, by changing the bed height from 50 to 150 mm (see Fig. 23a), drying costs increase; additionally, when the bed height is raised from 150 to 250 mm, drying costs decrease. This is an interesting result caused by the largest investment and operational costs for the bed height of 150 mm (see Attch. 1 of IV). Apparently, the average heat and mass transfer for the low bed height is so good that the ratio τ/z (see Eq. 5) becomes smaller for the bed height of 50 mm than for the bed height of 150 mm. This could explain the reason for drying costs becoming smaller for the bed height of 50 mm. For the bed height of 250 mm, this effect no longer exists. In [47], the ratio τ/z has also been studied. However, Holmberg and Ahtila found that despite the final moisture content of the wood particles, the ratio seems to decrease when the bed height increases. This result clearly differs from the results presented in this dissertation. This indicates that the bed material has influence on this ratio.

In conclusion, the drying costs of bark behave in the same way as the drying costs of woodchips; when the temperature is constant, the highest bed height yields the lowest drying costs. However, it can be stated that the birch bark is formed less than bark of spruce or pine in the Finnish wood industry. It can be assumed that the drying curves for spruce or pine bark can be slightly different and their drying cost evaluations might differ.

Soot sludge and sawdust mixture

Figure 24a shows the effect of bed height, air temperature and change of moisture content for the relative drying costs of sludge and sawdust mixture. Figure 24b shows the effect of different air velocities on relative drying costs.

Figure 24. The relative drying costs for soot sludge and sawdust mix. a) The effect of drying air temperature and the bed height on the drying costs, b) the effect of air velocities on the drying costs). In Fig. 24a, air velocity is 0.75 m/s in all cases. In Fig. 24b, bed height is 200 mm in both cases. [IV]

Figure 24a shows that increasing the temperature decreases drying costs. Generally, the drying phenomenon is limited by the average mass and heat transfer properties of the material bed as well as the water absorption capacity of air. The moister the material bed, the more important is the good water absorption capacity of the air, which enhances the removal of free water from the material bed. Increasing the air temperature enhances the water absorption capacity of air, which explains lower drying costs. However, increasing the air temperature from 80 to 100 °C does not have any remarkable effect on drying costs.

Fig. 24a also shows that increasing the bed height causes slightly higher drying costs (see the values in brackets), which differs from the results of woodchips and bark. This indicates that materials containing great amounts of free water (i.e. the initial moisture content is extremely high) have an optimal bed height which is lower compared to material with clearly lower initial moisture content.

Water absorption capacity of air can also be enhanced by raising the air velocity. However, in the case of the soot sludge and sawdust mix, higher air velocity does not decrease drying costs (see Fig. 24b). Attachment 1 of IV reveals that the drying times for both velocities are almost the same (same investment costs), but annual operational costs become higher for the velocity of 0.9 m/s. Drying costs behave in the same way when the air velocity is changed in the case of woodchips (see Fig. 22b).

For the soot sludge and sawdust mixture the investment costs might be slightly higher than those given by Eq. (4) because it does not take into account the mixing machine needed in the process.

4.3.2.2 Economic sensitivity analysis

Rate of interest

A sensitivity analysis with different rates of interest (5, 8, and 11 %) was conducted for woodchips and bark, as well as for the mix of soot sludge and sawdust in IV. The other values remained exactly the same as in the calculations

presented in the base cases. However, the rate of interest did not influence the relative drying costs of bark and the mix of soot sludge and sawdust when the drying air temperature, bed height, and air velocity (in the case of the sludge) were changed.

The only deviation noticed was when the influence of the rate of interest was studied on the relative drying costs of woodchips drying. Nevertheless, this slight derivation did not affect the main results of the case: the highest drying costs are gained with a bed height of 100 mm and air temperature of 50 °C and the lowest costs are gained with a bed height of 500 mm and air temperature of 90 °C as in the base case.

The relative drying costs of woodchip drying were also evaluated in IV when different air velocities and rates of interest were applied. The drying costs were at their lowest with an air velocity of 0.5 m/s, regardless of the rate of interest and the change of moisture content similar to the base case.

Prices of heat and electricity

A sensitivity analysis was conducted for the waste materials in IV for which the price of heat was changed for every drying air temperature (1, 5, 10, and 15 e/MWh), and one for which the price of electricity was changed (20, 30, 40, and 50 e/MWh). The heat price remained constant in the base cases of IV, despite the differences in the drying air temperatures, due to the heat pricing depending on the situation (see Section 2.2.3). To allow for the influence of the temperature on the heat price, the price of heat was changed for every drying air temperature in the sensitivity analysis. Attachment 2 of IV presents the results. The results show that the price of the heat has a significant impact on the economics of drying. In all cases, the lowest drying costs are achieved by the lowest air temperature, and drying costs become highest with the highest air temperature. For a given bed height, lowest drying costs may be about 30–75 % smaller than the highest ones depending on the material and the moisture content change. Thus, this result supports the idea of using low temperature heat for the drying if its price is clearly lower than the price of high temperature heat sources. Although the dryer size increases as the air temperature decreases, it seems to be more profitable to have a larger dryer due to its operational costs being smaller. This supports the statement in [12] on operational costs (see page 13).

The sensitivity analysis with different prices for electricity was also similarly performed as the one for the rate of interest. The results revealed that the electricity price does not influence the relative drying costs; furthermore, the results are similar to those in the base cases.

4.3.2.3 Block studies

The option of employing blocks with different air temperatures or velocities from the perspective of drying costs was discussed in IV. Figure 25 shows the analysed alternatives. Four alternatives for the manner of setting the air temperatures in the blocks were examined (see Fig. 25a). In alternative A.1, air with the higher air temperature (t_i) flows through the first block; and in the second block, air with the lower air temperature flows through the material bed

(t_2). In alternatives A.2 and A.3, the air temperature is the same in both blocks (t_2 or t_1 , respectively). In alternative A.4, air with the lower air temperature (t_2) flows through the first block; and in the second block, air with the higher temperature flows through (t_1).

Furthermore, three alternatives describing the air velocities set in the blocks were evaluated (see Fig. 25b). In alternative B.1, air with the higher air velocity (v_1) flows through the first block; and in the second block, air with the lower air velocity flows through the bed (v_2). In alternatives B.2 and B.3, air velocity is the same in both blocks (v_2 or v_1 , respectively).

A) BLOCK EVALUATION: Temperature of drying air, $t_1 > t_2$, v is constant

B) BLOCK EVALUATION: Velocity of drying air, $v_1 > v_2$, t is constant

Figure 25. Block evaluations: a) block evaluations in which temperature of drying air varies, b) block evaluations in which velocity of drying air varies. [IV]

All block evaluations were conducted so that the first block ended when 50% of the moisture content change had been removed. For example, when drying woodchips from the initial moisture content of $1.3 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (u_i) to the final moisture content of $0.2 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (u_2), the moisture content of material after the first block was $0.75 \text{ kg}_{\text{H}_2\text{O}}/\text{kg}_{\text{db}}$ (u_1). The block study was only implemented for woodchips as well as the mix of soot sludge and sawdust as in previous experiments [I, II] of these materials different air velocities were utilized. The initial values in the block calculations were the same as in the bases cases (Table 17).

Figure 26 shows the results of the block study in which the air temperature varied when woodchips were dried from the initial moisture content of 1.3 to

0.75, and lastly, to a final moisture content of 0.2 kg_{H₂O}/kg_{db}. The values in Fig. 26 represent relative drying costs. Number 1 in Fig. 26 always represents the highest drying costs for a given temperature combination with other costs in this line being compared to this value.

Alternative A.1 $t_1 - t_2$	Alternative A.2 $t_2 - t_2$	Alternative A.3 $t_1 - t_1$	Alternative A.4 $t_2 - t_1$
90°C - 30°C	30°C - 30°C	90°C - 90°C	30°C - 90°C
0.9	1	0.7	0.8
90°C - 50°C	50°C - 50°C	90°C - 90°C	50°C - 90°C
0.9	1	0.6	0.7
70°C - 30°C	30°C - 30°C	70°C - 70°C	30°C - 70°C
1 (0.96)	1	0.9 (0.88)	1 (0.95)
70°C - 50°C	50°C - 50°C	70°C - 70°C	50°C - 70°C
1 (0.97)	1	0.7	0.8

Figure 26. Block evaluations for woodchips when two blocks with varying air temperatures were used. Drying of woodchips from the initial moisture content of 1.3 to 0.75, and lastly to a final moisture content of 0.2 kg_{H₂O}/kg_{db}. The bed height was 300 mm and air velocity 0.7 m/s in all calculations. [IV]

Figure 26 shows that the drying costs are highest when the lowest air temperature is used in both blocks (see the second column of Fig. 26). The lowest drying costs were achieved by using the highest air temperature in both blocks. The results in Fig. 26 also show that the drying costs became smaller when the first block used lower drying temperature than the second block ($t_2 - t_1$) (A.1 vs. A.4). This indicates that the drying process is more controlled by the heat and mass transfer over the bed than the water absorption capacity of air.

The block study with varying air temperatures (30, 50, 70, and 90 °C) was also conducted for woodchips when they were dried from the initial moisture content of 1.3 to 0.95, and lastly, to a final moisture content of 0.6 kg_{H₂O}/kg_{db}. The results were similar in the case where the change of the moisture content was 1.3–0.2 (see Fig. 26). For the soot sludge and sawdust mix (5.3 to 3, and lastly, to 0.7 kg_{H₂O}/kg_{db}), the relative drying costs also behaved similarly as in Fig. 26. Based on these results, material does not seem to influence the way heat sources should be optimally used in a belt dryer.

These results supports the result in [45]. Holmberg et al. [45] conclude that it might be possible to divide the dryer into more than one blocks in some cases when drying costs are optimised.

When materials are dried, the availability of heat sources at different temperatures may be limited. In that case, it may be reasonable to use different air temperatures in successive drying blocks. Based on the results, it seems to be more economic to use lower air temperatures in the first blocks when the material is moist, and higher air temperatures in the last blocks when the material is drier. Alternatively, air can be heated stepwise to the higher temperature using heat sources at different temperatures in heating (see e.g. [47]). If different air temperatures are used in the blocks, it is reasonable to analyse the way heat sources should be used from the viewpoint of drying costs.

Figure 27 presents the results for drying woodchips as well as the mixture of soot sludge and sawdust with different air velocities.

Alternative B.1 $v_1 - v_2$	Alternative B.2 $v_2 - v_2$	Alternative B.3 $v_1 - v_1$
0.7 m/s - 0.3 m/s	0.3 m/s - 0.3 m/s	0.7 m/s - 0.7 m/s
0.93	0.95	1

a) Woodchips:

change of moisture content 1.3 - 0.75 - 0.2 kg_{H₂O} / kg_{db}

The bed height 300 mm, air temperature 70 °C in all calculations.

Alternative B.1 $v_1 - v_2$	Alternative B.2 $v_2 - v_2$	Alternative B.3 $v_1 - v_1$
0.7 m/s - 0.3 m/s	0.3 m/s - 0.3 m/s	0.7 m/s - 0.7 m/s
0.95	1	0.87

b) Woodchips:

change of moisture content 1.3 - 0.95 - 0.6 kg_{H₂O} / kg_{db}

The bed height 300 mm, air temperature 70 °C in all calculations.

Alternative B.1 $v_1 - v_2$	Alternative B.2 $v_2 - v_2$	Alternative B.3 $v_1 - v_1$
0.9 m/s - 0.75 m/s	0.75 m/s - 0.75 m/s	0.9 m/s - 0.9 m/s
0.92	0.91	1

c) Soot sludge and sawdust mix:

change of moisture content 5.3 - 3.0 - 0.7 kg_{H₂O} / kg_{db}

The bed height 200 mm, air temperature 80 °C in all calculations.

Figure 27. The block study in which air velocity of drying air varied. [IV]

The results in Fig. 27 show that the change of moisture content impacts drying costs when the air velocities are changed in the drying blocks. Nevertheless, the differences between drying costs in Fig. 27 are quite small. It is difficult to draw any general conclusion on optimal air velocities in different blocks. This result supports the assumption made in the base cases of IV that the air velocity through the material bed is the same over the whole dryer.

5. Conclusions

The utilization of excess heat from industry in the drying of waste streams, which must be disposed of by combustion, enhances the energy efficiency of industrial processes and reduces global emissions. In many cases, the moisture content of a waste stream is high. The waste streams are combustible with lower moisture contents. The drying of a waste stream improves the lower heating value of a stream. A suitable dryer for wastes is a belt dryer when lower air temperatures are used. However, the total drying costs may vary considerably

depending on how different design parameters are chosen, such as bed height, air temperature, and air velocity. In addition, factors that affect the drying costs include the heat source (temperature level, price) and the cost of electricity.

The main objective of this dissertation was to investigate the effect of the design parameters of the belt dryer on the total drying costs when different waste streams are dried using low temperature excess heat (below 100 – 120 °C). The influence of the following design parameters were studied: bed height, drying air temperature, air velocity and initial/final moisture contents of the material. This dissertation studied the drying of three waste streams: woodchips, bark and the mix of the soot sludge and sawdust. The waste streams examined were real industrial waste.

In order to produce a techno-economic evaluation for belt drying, the drying kinetics of the waste materials had to be discovered. A series of experimental test were conducted for woodchips, and the mix of the soot sludge and sawdust to study their drying kinetics in fixed beds (Publications I-III). Additionally, drying curves for bark produced by Holmberg et al. [21] were used in the techno-economic evaluation. A generic model was formed in Publ. IV in order to determine the drying costs of upgrading of the previously mentioned waste streams to fuel by means of drying in a belt dryer. The evaluation was completed using the annuity method.

Based on the key observations of this dissertation, the following main conclusions can be drawn on drying kinetics:

The air temperature has a great impact on the drying time for all materials; the lower the drying air temperature, the longer the drying time needed. For example, for woodchips, the drying time is almost two times longer when the temperature decreases from 70 to 50 °C (bed height 300 mm).

Increasing the air velocity from 0.3 to 0.5 m/s has a remarkable influence on the drying time of woodchips (3.7 h versus 2.2 h, air temperature 70 °C and bed height 300 mm). In contrast, the difference in drying times between air velocities of 0.5 and 0.7 m/s is quite small (2.2 h and 1.9 h, respectively). This indicates that it would be reasonable to use an air velocity of at least 0.5 m/s. In the case of soot sludge, there is no reason to use an air velocity higher than 0.75 m/s, because drying times for the desired final moisture content of 0.7 kg_{H₂O}/kg_{db} are at the same magnitude when using air velocities of 0.75 and 0.9 m/s. These results indicate that there is an optimum value for air velocity, after which an increase of the air velocity will no longer remarkably decrease the drying time. Furthermore, using higher air velocities increases the pressure drop over the bed and the heat consumption.

The most optimal volume ratio for the soot sludge and sawdust is 50%_{soot sludge} and 50%_{sawdust}. Results showed that when the share of the soot sludge was higher, the mixture did not dry properly and the drying time was considerably lengthened. When the volume ratio was 70%_{soot sludge} and 30%_{sawdust}, the drying time became approximately over 4 times longer compared to the ratio of 50%:50%. The tests also showed that the material dried quite evenly, even though the bed height was different when the mixture ratio of 50%: 50% was used. On the basis of the results, the most optimal bed height is 200 mm.

Publs. I and II noted that the pressure drop over the bed constantly decreases as the drying proceeds. The behaviour of the pressure drop probably results from the slight movement of the material in the bed as it dries, and because the void volume increases in the bed. Additionally, as expected, the pressure drop increases when the bed height of the material becomes higher.

The linear characteristic drying curves for the soot sludge and sawdust mixture can be used to extrapolate drying data from specific drying conditions to another when the bed height is 200 or 300 mm and air temperature 40 – 100 °C (or little higher air temperature, too). The characteristic drying curves estimate drying times with a good accuracy.

The average values of the volumetric heat transfer coefficient were defined for the soot sludge and sawdust mixture in Publ. III. The average values for the term αa for the bed heights of 200 and 300 mm are 16443 and 11745 W/(m³K), respectively. These values can be used to calculate the constant drying rate for the sludge and sawdust mix in a fixed bed when the inlet air temperature is ≤ 100 °C. The average αa value can also be used to estimate the constant drying rate although the inlet temperature is slightly higher than 100 °C if more accurate data is not available.

On the basis of the results, the following main conclusions can be drawn on the economic calculations:

The results of Publ. III indicated that it is economical to dry soot sludge with sawdust in order to reduce the oil consumption in a boiler.

The results of the Publ. IV for the woodchips and bark showed that the highest bed height (bark: 250 mm, woodchips: 500 mm) and the highest air temperature (bark: 110 °C; woodchips: 90 °C) yield the lowest drying costs, regardless of the change of moisture content, when the price of the heat has been kept constant for every air temperature. In the case of the soot sludge and sawdust mix, the lowest drying costs are gained with the highest air temperature. However, raising the bed height from 200 to 300 mm slightly increases drying costs. This indicates that there is an optimal bed height in a belt dryer, which depends on the material and its initial moisture content. The higher the moisture content, the lower the optimal bed height seems to be. Increasing the air velocity does not necessarily enhance the drying rate after a certain value, but may increase the operational costs of the dryer.

The sensitivity analyses indicated that the rate of interest or the electricity price does not influence drying costs. The heat price affects the optimal design parameters. The results of the heat price sensitivity analysis revealed that the lowest drying costs are achieved using the lowest air temperature when the heat price depends on its temperature level. The results support the idea of using low temperature heat sources in drying if they are clearly cheaper than high temperature heat sources. However, if the price for all heat sources at different temperatures is at the same magnitude, the high temperature heat sources should be favoured.

When materials are dried, the availability of heat sources at different temperatures may be limited. In that case, it may be reasonable to use different air temperatures in successive drying blocks. On the basis of block studies, it

seems to be more economic to use lower air temperatures in the first blocks when material is moist, and higher air temperatures in the last blocks when the moisture content of material is lower if the heat price is the same for both temperature levels. When air velocities were changed in the blocks, the optimal combination seemed to be case-dependent. Nevertheless, the changes between the drying costs were not remarkable when these velocities were changed. On the basis of these results, air velocity should be the same over the entire dryer.

As Adamiec, states in [96] differences of various waste sludges mean that their drying kinetics may notably differ from each other. Due to this, it is difficult to apply the drying kinetics of a certain sludge to another sludge with different origin. This means that experimental tests for soot sludge and sawdust mixture and the drying kinetics studies of this dissertation give highly valuable new information. However, it is worth of mentioning that soot sludge and sawdust mixture drying in a large-scale industrial dryer may have some challenges, for example, dusting can occur when dry mixture comes out of the dryer. For a future work, it would be interesting to research how drying is accomplished with a real drier and the effect that dried fuel has on boiler fuel consumption and other operations.

In several cases, the energy efficiency of a dryer is enhanced, for example, by using the heat recovery of the exhaust air of a dryer or partial recirculation of the drying air, which affects the overall drying costs and must be considered in the design. However, in this dissertation the enhancement methods were not included to the study.

In general, the results showed that it is important to pay attention to the main design parameters to optimize total drying costs. For example, if an overly low bed height is used in woodchips or bark drying, the total drying costs might be dozens of per cent higher than in the most economic case. This indicates that it is extremely important to select the correct design parameters for a belt dryer. This dissertation also reveals that the price of the heat has a significant impact on the economics of drying.

The results of this dissertation have justified reasonably well Mujumdar's assertion mentioned on page 13 in the Introduction. Incremental improvements in dryer design can often mean substantial savings over the life of a dryer.

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ISBN 978-952-60-8833-4 (printed)
ISBN 978-952-60-8834-1 (pdf)
ISSN 1799-4934 (printed)
ISSN 1799-4942 (pdf)

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