TKK Dissertations 63
Espoo 2007

BIOFUEL DRYING AS A CONCEPT TO IMPROVE THE ENERGY EFFICIENCY OF AN INDUSTRIAL CHP PLANT

Doctoral Dissertation

Henrik Holmberg

Helsinki University of Technology
Department of Mechanical Engineering
Laboratory of Energy Economics and Power Plant Engineering

TKK Dissertations 63
Espoo 2007

BIOFUEL DRYING AS A CONCEPT TO IMPROVE THE ENERGY EFFICIENCY OF AN INDUSTRIAL CHP PLANT

Doctoral Dissertation

Henrik Holmberg

Dissertation for the degree of Doctor of Science in Technology to be presented with due permission of the Department of Mechanical Engineering for public examination and debate in Auditorium K216 at Helsinki University of Technology (Espoo, Finland) on the 13th of April, 2007, at 12 noon.

Helsinki University of Technology
Department of Mechanical Engineering
Laboratory of Energy Economics and Power Plant Engineering

Teknillinen korkeakoulu
Konetekniikan osasto
Energiatalouden ja voimalaitostekniikan laboratorio

Distribution:
Helsinki University of Technology
Department of Mechanical Engineering
Laboratory of Energy Economics and Power Plant Engineering
P.O. Box 4100
FI - 02015 TKK
FINLAND
URL: http://www.tkk.fi/Yksikot/Energiatalous/
Tel. +358-9-451 3621
Fax +358-9-451 3674
E-mail: helena.eklund@tkk.fi

ISBN 978-951-22-8648-5
ISBN 978-951-22-8649-2 (PDF)
ISSN 1795-2239
ISSN 1795-4584 (PDF)
URL: http://lib.tkk.fi/Diss/2007/isbn9789512286492/

TKK-DISS-2270

Otamedia Oy
Espoo 2007

HELSINKI UNIVERSITY OF TECHNOLOGY P. O. BOX 1000, FI-02015 TKK http://www.tkk.fi	ABSTRACT OF DOCTORAL DISSERTATION
Author Henrik Holmberg	
Name of the dissertation Biofuel drying as a concept to improve the energy efficiency of an industrial CHP plant.	
Date of manuscript August 9, 2006	Date of the dissertation April 13, 2007
☐ Monograph	**X** Article dissertation (summary + original articles)
Department Mechanical Engineering Laboratory Energy Economics and Power Plant Engineering Field of research Industrial Energy Engineering Opponent(s) Prof. Thore Berntsson, Dr. Pat McKeough Supervisor Prof. Pekka Ahtila (Instructor) Prof. Markku Lampinen	

Abstract

This thesis presents the results of a research project in which the enhancement of the CHP production by means of biofuel drying at an integrated pulp and paper mill has been explored. Instead of high temperature flue gas drying the use of low-temperature secondary heat as drying energy has been studied. The drying system considered in this thesis is classified as follows: Drying medium: Air Heat supply into the material: Convection Transport mechanism of the material inside the dryer: Conveyor Method to improve energy efficiency in the drying system: Multi-stage drying. In addition to secondary heat, the option of using steam as a drying energy has been included in the calculation models.

An optimization model for analyzing the integration of a multi-stage drying system into the CHP process has been developed. A simulation model based on the optimization model has been created to analyze the operation of the dryer under variable drying conditions. The simulation model can also be applied to the model-based control of the final fuel moisture content if there is a need to control it. Both models have been applied to a case study. Drying as a physical phenomenon has been studied experimentally in a small fixed bed dryer, and guidelines for dimensioning a continuous conveyor dryer in the case of multi-stage drying are presented. The energy efficiency of drying has also been analyzed using two evaluation methods: specific heat consumption and the irreversibility rate.

According to case study results, an optimally designed dryer uses only secondary heat, not steam, as a heat source, and earnings stem from decreased marginal fuel consumption, not increased power generation. Simulation calculations show that there are no economic grounds for control of the final fuel moisture content by adjusting heat inputs into the dryer. To control the final fuel moisture content the use of homogenous, and not just dry, fuel must have some positive influences on the operation of the boiler and/or power plant. Experimental tests show that critical moisture content is high for woody-based fuels and diffusion-based drying models must be used to determine drying times theoretically in a fixed bed. If the energy used in drying can be converted to mechanical work, the irreversibility rate is a more comprehensive method of comparing energy efficiency between different drying processes than specific heat consumption.

Keywords Biofuel drying, CHP production, Secondary heat	
ISBN (printed) 978-951-22-8648-5	ISSN (printed) 1795-2239
ISBN (pdf) 978-951-22-8649-2	ISSN (pdf) 1795- 4584
ISBN (others)	Number of pages 63 p. + app. 62 p.
Publisher Helsinki University of Technology, Dept. of Mech. Eng., Lab. of Energy economics and Power Plant Eng.	
Print distribution Laboratory of Energy Economics and Power Plant Engineering	
X The dissertation can be read at http://lib.tkk.fi/Diss/2007/isbn9789512286492/	

TEKNILLINEN KORKEAKOULU PL 1000, 02015 TKK http://www.tkk.fi	VÄITÖSKIRJAN TIIVISTELMÄ
Tekijä Henrik Holmberg	
Väitöskirjan nimi Biopolttoaineen kuivauksen mahdollisuudet parantaa energiatehokkuutta metsäteollisuuden CHP-tuotannossa	
Käsikirjoituksen jättämispäivämäärä 9.8.2006	Väitöstilaisuuden ajankohta 13.4.2007
☐ Monografia	X Yhdistelmäväitöskirja (yhteenveto + erillisartikkelit)

Osasto	Konetekniikka
Laboratorio	Energiatalous ja voimalaitostekniikka
Tutkimusala	Teollisuuden energiatekniikka
Vastaväittäjä(t)	Prof. Thore Berntsson, Tohtori Pat McKeough
Työn valvoja	Prof. Pekka Ahtila
(Työn ohjaaja)	Prof. Markku Lampinen

Tiivistelmä

Väitöstyössä esitetään tuloksia ja johtopäätöksiä tutkimusprojektista, jossa on tutkittu yhdistetyn sähkön ja lämmöntuotannon (CHP-tuotanto) energiatehokkuuden parantamista biopolttoaineen kuivauksen avulla integroidussa sellu- ja paperitehtaassa. Perinteisen savukaasukuivauksen sijasta tutkimuksessa on analysoitu matalalämpöisen sekundäärilämmön soveltuvuutta biopolttoaineen kuivaukseen. Laskentamallit on muodostettu kuivaussysteemille, jossa kuivauskaasuna toimii ilma, lämmönsiirto kuivauskaasusta materiaaliin tapahtuu konvektiolla, materiaalin siirto kuivurissa toteutetaan jatkuvatoimisella kuljettimella ja kuivurin energiatehokkuutta voidaan tarvittaessa parantaa ilman vaiheistuksen avulla (ns. monivaihekuivaus). Sekundäärilämmön lisäksi laskentamalleissa huomioidaan mahdollisuus käyttää vastapaine- ja/tai väliottohöyryä kuivausilman lämmityksessä.

Työssä on muodostettu optimointimalli monivaihekuivurin liittämiseksi CHP-prosessiin sekä simulointimalli kuivurin toiminnan analysoimiseksi muuttuvissa kuivausolosuhteissa. Simulointimallia voidaan tarvittaessa käyttää polttoaineen loppukosteuden mallipohjaisessa säätämisessä. Sekä optimointi- että simulointimallia on sovellettu esimerkkitapauksissa, joiden lähtöarvot ovat peräisin eräältä suomalaiselta sellu- ja paperitehtaalta. Kuivumista fysikaalisena ilmiönä on tutkittu kokeellisesti pienessä laboratoriokokoluokan kiintopetireaktorissa, ja koetuloksiin perustuen esitetään laskentaperiaatteet kuivurin mitoittamiseksi monivaihekuivauksessa. Työssä on lisäksi analysoitu kuivurin energiatehokkuutta sekä ominaisenergiankulutuksen että exergia-menetelmän perusteella.

Tulosten perusteella optimaalisesti mitoitettu kuivuri käyttää useimmissa tapauksissa lämmönlähteenä ainoastaan sekundäärilämpöä ja kuivauksen tuotot muodostuvat kokonaisuudessaan marginaalipolttoineen kulutuksen vähenemisestä, eivätkä lisääntyneestä sähköntuotannosta. Simulointilaskelmissa tehtyjen oletusten perusteella polttoaineen loppukosteutta ei ole perusteltua säätää vaan sen kannattaa antaa mieluummin vaihdella. Loppukosteuden säätö on perusteltua, jos tasalaatuisen, ei siis pelkästään kuivatun, polttoaineen käytöllä on joitain positiivisia vaikutuksia kattilan ja/tai voimalaitoksen toimintaan. Näitä mahdollisia vaikutuksia ei ole työssä analysoitu. Kokeelliset mittaukset osoittavat, että kriittinen kosteuspitoisuus on korkea puuperäisille materiaaleille ja kuivumisaikojen teoreettisessa laskennassa on käytettävä diffuusioteoriaan perustuvia kuivumismalleja. Jos kuivaukseen käytettävää energiaa voidaan muuttaa mekaaniseksi työksi alueella, jossa kuivuri sijaitsee, pitäisi erilaisten kuivureiden energiatehokkuutta verrata mieluummin exergia-menetelmään perustuen kuin ominaisenergiankulutusten perusteella

Asiasanat Biopolttoaineen kuivaus, CHP-tuotanto, sekundäärilämpö	
ISBN (painettu) 978-951-22-8648-5	ISSN (painettu) 1795-2239
ISBN (pdf) 978-951-22-8649-2	ISSN (pdf) 1795- 4584
ISBN (muut)	Sivumäärä 63 s. + julkaisut 62 s.
Julkaisija Teknillinen korkeakoulu, Konetekniikan osasto, Energiatalouden ja voimalaitostekniikan laboratorio	
Painetun väitöskirjan jakelu Energiatalouden ja voimalaitostekniikan laboratorio	
X Luettavissa verkossa osoitteessa http://lib.tkk.fi/Diss/2007/isbn9789512286492/	

Preface

This work has been done at the Laboratory of Energy Economic and Power Plant Engineering (EVO), Helsinki University of Technology. The study has been funded by Pohjolan Voima Oy, Stora Enso Oyj, Vapo Oy and National Technology Agency of Finland (Tekes). All financers are gratefully acknowledged for making this work possible.

I would like to express my gratitude to my supervising professor Pekka Ahtila for his guidance, comments and several good discussions during this work. In some cases, these discussions have had nothing to do with drying or energy efficiency but they have also been important and pleasant. Professor Markku Lampinen has been my instructor in this work and I want to thank him for useful comments and also for a good basic teaching that his has given on his courses. The working atmosphere at the laboratory of EVO has been pleasant and encouraging during these years. Thanks to all current and ex-colleagues for that. Special thanks to Ilkka H. and Mari for several valuable comments to my scientific output and for helping me in different kinds of practical problems.

Finally, I want to thank my family members and friends. Even if most of you have had nothing to with the thesis itself it would have been much harder to complete this long process without you.

Helsinki, February 2007
Henrik Holmberg

Contents

List of papers

The thesis is based on the following six papers

I Holmberg H, Ahtila P. Drying Phenomena in a Fixed Bed Under Biofuel Multi Stage Drying. In: Oliveira A, Afonso C, Riffat S, editors. Proceedings of the 1st International Conference of Sustainable Energy Technologies, Porto, Portugal; June 12-14, 2002. p. EES1 6-11.

II Holmberg H, Ahtila P. Comparison of drying costs in biofuel drying between multi-stage and single-stage drying. Biomass&Bioenergy 2004; 26: 515-530.

III Holmberg H, Ahtila P. Optimization of the bark drying process in combined heat and power plant process of pulp and paper mill. In: Odilio AF, Eikevik TM, Strommen I, editors. Proceedings of the 3rd Nordic Drying Conference, Karlstad, Sweden; June 15-17, 2005.

IV Holmberg H, Ahtila P. Adjusting of temperature levels in multi stage drying system by means of outlet air measurements. In :Oliveira A, Afonso C, Riffat S, editors. Proceedings of the 1st International Conference of Sustainable Energy Technologies, Porto, Portugal; 12-14 June, 2002. p. EES2 1-5.

V Holmberg H, Ahtila P. Simulation model for the model-based control of a biofuel dryer at an industrial combined heat and power plant. Drying Technology 2006; 24:1547-1557.

VI Holmberg H, Ahtila P. Evaluation of energy efficiency in biofuel drying by means of energy and exergy analyses. Applied Thermal Engineering 2005; 25: 3115-3128.

All papers are independent research carried out and written by the author. Professor Ahtila has contributed to the theoretical content of the papers. He has also commented all papers.

Nomenclatures

A	Cross-sectional area of the dryer [m^2]
A_e	Evaporation surface [A_e]
b	Energy price [€/MWh]
C	Cost [€]
c_p	Specific heat capacity [J/kgK]
D_{AB}	Diffusion coefficient of vapour-air mixture [m^2/s]
D_{eff}	Effective diameter [m]
E	Exergy [W]
q	Specific heat consumption [J/kg]
L	Length of conveyor [m]
Le	Lewis number [-]
I	Irreversibility rate [W]
l_v	Vaporisation heat of water [J/kg]
$\dot{M}_{dm}$	Dry mass flow of fuel [kg/s]
M_v	Molecular mass of water [kg/mol]
$\dot{m}_{da}$	Dry mass flow of air [kg/s]
$\dot{m}_e$	Evaporation rate [kg/s]
m	Mass [kg]
NI	Net income [€]
NPV	Net present value [€]
Nu	Nusselt number ($\alpha D_{eff}/\lambda$) [-]
p_o	Total pressure [Pa]
p_v	Partial pressure of vapour [Pa]
p_v'	Saturated vapour pressure [Pa]
Pr	Prandtl number ($\eta c_p/\lambda$) [-]
Re	Reynolds number (vD_{eff}/ν) [-]
s	Entropy [J/kgK]
T_o	Temperature of surrounding [K]
T,t	Temperature [K,^{o}C]
u	Fuel moisture content [kg/kg_{dm}]
V	Volume [m^3]
W	Work [W]
v	Velocity [m/s]
x	Air moisture content [kg/kg_{da}]
Z	Bed thickness [m]

Greeks

α	Heat transfer coefficient [W/m^2K]
ε	Volume fraction of air in the fuel bed [-]
Φ	Heat flux, heat consumption [W]
μ	Dynamic viscosity [kg/ms]

λ	Heat conductivity [W/mK]
ρ	Density [kg/m^3]
ν	Kinematic viscosity [m^2/s]
τ_u	Drying time [s]
τ_{op}	Operating hours [h]
$\Delta\tau$	Length of time interval [s]

Subscripts

a	Air
bs	Backpressure steam
c	Conveyor
d	Dryer
da	Dry air
db	Dry bulb
dm	Dry mass
e	Electricity
es	Extraction steam
f	Fuel
in	Inlet, initial
mf	Marginal fuel
out	Outlet, final
s	Surface
sh	Secondary heat
v	Vapour
wb	Wet bulb

1 Introduction

1.1 Biofuels in the forest industry

Classification and properties of biofuels

Material of biological origin excluding material embedded in geological formations and transformed to fossil is called biomass [1]. Biomass can be further divided into woody and herbaceous biomasses. Fuels produced directly or indirectly from these biomasses are called biofuels [1].

All biofuels consumed by the Finnish forest industry are obtained from woody biomasses and can be divided into three main groups: black liquor, sludge, and solid biofuels. Figure 1 illustrates sources from which biofuels are obtained in the forest industry. Black liquor is the most important biofuel in Finland [2]. Sludge streams are usually small compared with other biofuel streams at pulp and paper mills. The third group, solid biofuels, is composed of bark, forest residues, sawdust, cutter shavings, and recycled wood-based waste (e.g. construction wood, crushed or chipped used wood, used paper and board) [3,4]. In 2004, the Finnish forest industry consumed approximately 60,000 TJ solid biofuels as mill fuels [2,5,6]. This thesis focuses on the drying of solid biofuels, and hereinafter the term biofuel always means this group. Figure 2 shows examples of the most common biofuels used by the forest industry in Finland.

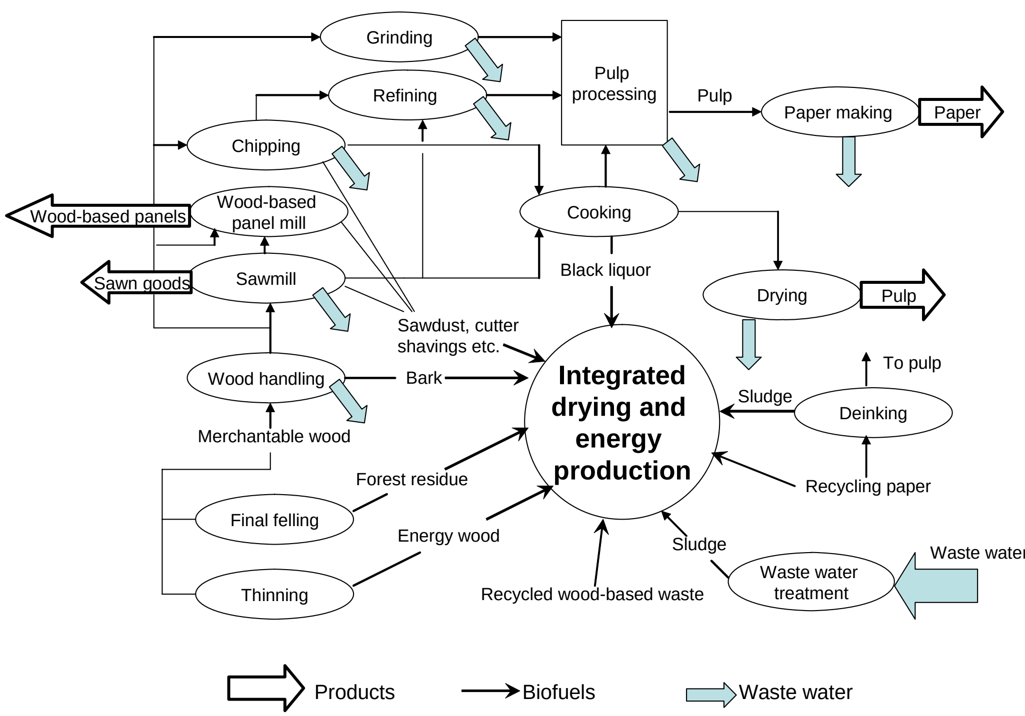


Figure 1. Sources of biofuels in the forest industry, reproduced from [4].

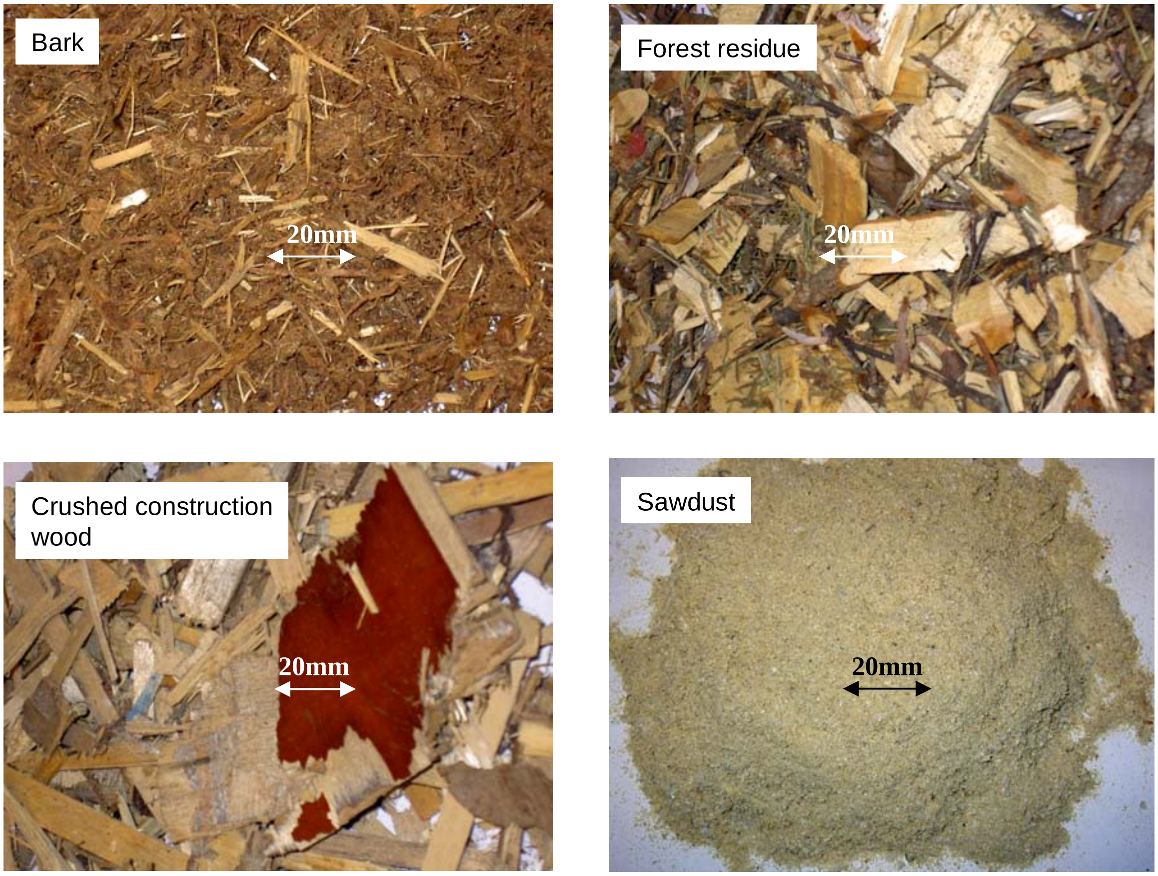


Figure 2. Examples of the most common biofuels used by the forest industry in Finland.

The average moisture content of bark is typically between 55-60% w.b. depending on the bark type [7]. In addition to the high average moisture content, the weather, season, and storage time may cause drastic deviations in the bark moisture. The moisture content of sawdust may vary from air-dry to even 70% w.b. [8]. On the other hand, cutter shavings usually have a low moisture content of 5 and 15 % w.b. [8]. The moisture content of the used wood-based products depends on its source. Usually, the material is relatively dry.

Forest residue consists of wood material that is left in forest after logging or thinning. Forest residue is crushed before combustion. The moisture content depends on the logistic chain. If the forest residue is delivered directly to the mill, the fuel has almost the same moisture content as the fresh wood. If the residue is allowed to dry after harvesting, the moisture content may decrease even to 20-30 % w.b. during the summer months [8].

Benefits of drying

All boilers used for the combustion of biofuels in Finland are fluidized bed boilers: either the bubbling fluidized bed boiler (BFB boiler) or the circulating fluidized bed boiler (CFB boiler). Fluidized bed boiler technology enables the combustion of moist fuels, and at present biofuels are not yet dried before combustion in the forest industry. Compared with moist biofuel the use of dry fuel would offer the following benefits:

- Improvement of the effective heating value
- Homogenization of the heating value resulting from the variation in fuel moisture content
- Smaller boiler dimensions
- Lower amounts of unburned solid and gaseous compounds from the boiler

The effective heating value (also known as the lower heating value) is directly proportional to the fuel moisture content. As a result of drying, energy input into the boiler may be increased without increasing the fuel input, or the fuel input into the boiler may be decreased to get the same energy input as in the case of moist fuel.

The basic functions of the combustion control and burner management systems are to maintain constant steam flow or pressure under varying loads through proper input of fuel and to maintain safe and efficient operation throughout the boiler's load range [9]. Compared with several other fuels (e.g. oil, natural gas and coil) the heating value of biofuel varies constantly as a result of varying moisture content. It is possible to control large changes in fuel quality in fluidized bed combustion, but such boilers set high requirements for the process control [10]. At the moment, the trend seems to be towards more advanced process control systems in the boiler [11]. However, drying can be seen as a conceivable method of decreasing the need to control the boiler caused by the varying fuel quality. The need to control the final fuel moisture content in the dryer is discussed in Chapter 4.3.

Boiler dimensions may be reduced if the boiler is designed for dry fuel. However, this requires that the operation of the dryer is sufficiently robust. If moist fuel cannot be dried all the time, a boiler designed for dry fuel may become a bottleneck in steam generation, or fossil support fuels must be used.

Generally, increasing moisture content means a more incomplete combustion. For example, increasing fuel moisture content means higher emissions of hydrocarbons due to an incomplete combustion [12]. However, it is difficult to obtain any unequivocal correlation between fuel moisture content and emissions, since the type of equipment and the mode of operation also affect the results [12].

1.2 Classification of biofuel dryers

Biofuel drying systems may be classified using three principal factors:

1. **Drying medium**
   Flue gas
   Air
   Steam
2. **Heat supply into the material**
   Convection (direct dryers)
   Conduction (indirect dryers)
   Combination of direct and indirect dryers
3. **Transport mechanism of the material inside the dryer**
   Rotary/drum dryers
   Conveyor dryers (also known as fixed bed dryers)
   Paddle/screw dryers
   Fluidized bed dryers
   Cascade dryers
   Pneumatic dryers

Further sub-classification to continuous or batchwise dryers is possible but usually unnecessary. Table 1 shows a short summary of the most common direct dryers used in biofuel drying based on references [12-21].

Table 1. Summary of the most common direct dryers used in biofuel drying [12-21]

Dryer type	Typical drying medium	Reported drying temperatures	Reported evaporation rates	Reported heat and electricity consumptions	Reported good sides [14,17,18,19]	Reported problems/bad sides [14,17,18,19]	Current suppliers
Drum dryers	Flue gas	200-600°C[13]	3.6-20t_{H2O}/h[14]	Heat 3.5-4.2 MJ/kg_{H2O}[16] Electricity 10-50 kWh/t_{dm}[16]	- Suitable for fuels with heterogeneous particle size - Robust - Low maintenance costs	- Dust and smell problems - Too coarse bark has caused blockage - Fire risk after the dryer and in shutdown	GEA, Torkapparater, DryCo
Conveyor dryers (also known as fixed bed dryers)	Air	30-150°C[14,18]	0.5-40t_{H2O}/h[14,18]		- Suitable for fuels with heterogeneous particle size - Suitable for low temperature drying with long residence times - Robust - Good controllability	- Large dryer dimensions - Fire risk inside the dryer	Swiss Combi, Bruks Klöckner, Mabarex, Andritz Fiber Drying
Cascade dryers	Flue gas	160-280°C[13]	0.8-7t_{H2O}/h[14]	Heat 5.8 MJ/kg_{H2O} (share of flue gas loss 2.5 MJ/kg_{H2O})[16] Electricity 10-15 kWh/t_{dm}[16]	- Suitable for fuels with heterogeneous particle size - Reasonable dryer dimensions - Robust in most cases	- Corrosion and erosion - Fire risk after the dryer and in shutdown - Long bark stripes have entangled around moving parts	Have been supplied by Bacho Industri. Current situation unclear.
Pneumatic dryers (e.g flash and mill dryers)	Flue gas Steam	- For flue gas dryers 150-700°C[13,15] - For steam dryers temperature depends on the pressure and degree of super heating usually t_{dry} > 150°C[14,19]	Flue gas dryers 10-26 t_{H2O}/h[14,19] Steam dryers 6-30t_{H2O}/h[14,19]	Heat (flue gas) 3.7 MJ/kg_{H2O}[16] Electricity (flue gas) 60-120 kWh/t_{dm} (in mill dryers share of grinding 25-40kWh/t_{dm})[14,16] Heat (steam) 400-1000 kJ/kg_{H2O}, when heat generated in the dryer is recovered[12,14,20]	- Small dryer dimensions - Flue gas dryers have been robust - Heat consumption is small for steam dryers, if heat generated in the dryer is recovered	- Not as suitable for large particles as other dryer types - Corrosion and especially erosion problems => high maintenance costs - Fire risk after the dryer and in shutdown - Leakages and fuel in- and output have caused problems in steam dryers	GEA DryCo Einco

Additional notes:
- Cascade dryer can be regarded as a kind of application of traditional fluidized bed dryer and pneumatic dryer, and have been quite widely used for biofuel drying, especially in Sweden.
- Fluidised bed dryers are not favoured in biofuel drying, because most biofuels fulfil unsatisfactorily criteria (see Ref. [21]) set on material suitable for fluidised bed drying.
- Paddle/screw dryers represent usually indirect dryers, and are quite rarely used for drying of solid biofuels. However, some applications of screw/paddle dryers used for sawdust drying are mentioned in [14].

1.3 Background of biofuel drying in the forest industry

Even though biofuel drying is not common before combustion at the moment, commercial biofuel dryers were used at pulp and paper mills in the 1970s and 1980s. The dominant combustion technique for biofuels at that time was grate firing. This type of boiler can handle fuels with varying moisture content, but ideally a moisture content of 30-40% w.b. should be used [22]. The main reason for dryer investments in the 1970s and the 80s was probably the high oil price resulting from two oil crises. In some cases, the moist fuel also decreased the boiler capacity so much that it became reasonable to install a dryer in combination with the boiler. In the 1970s and 1980s, all industrial dryers in the Finnish forest industry were direct flue gas dryers [13]. Flue gases were either taken directly from the boiler or generated in a separate flue gas burner [13]. According to references [13-15] the most common dryer types were drum dryer, cascade dryer and pneumatic dryer.

Since the 1970s fluidized bed boilers have replaced grate firing as a combustion technique [7]. Compared with grate firing, the fluidized bed boiler is a more suitable combustion technique for moist biofuels. The moisture content must be higher than approximately 62-65% w.b. before stable combustion becomes difficult to maintain and it becomes necessary to support the firing with some fossil fuel [7]. However, the use of moist fuel decreases the energy efficiency of the power plant, and one may ask why the existing flue gas dryers were not integrated with fluidized bed boilers. At least, the following reasons may be listed:

- Bad operating experiences
- Environmental considerations
- Economic factors

Biofuel is a burnable material with a heterogeneous particle size. It is also typical that the biofuel flow contains stones, sand and other inappropriate things. It is obvious that the properties of biofuel set tough requirements for the operation of the dryer. For some dryer types (e.g. pneumatic dryers) the maintenance cost may also be high. It is presumable that the operational experiences of flue gas dryers have not always been satisfactory and bad experiences have supported the decision to dispense with the dryer. It is still one of the main prerequisites for drying that the dryer is a robust technique. The advantages and drawbacks of different dryers are listed in Table 1.

All types of wood material contain volatile organic material that may be emitted together with the water vapour [12]. There are presently legal restrictions on the amounts that may be released. The emissions of biomass drying are greatly affected by the drying temperature, when it exceeds 100^{o}C [23]. Below 100^{o}C emissions are reported to be low [23, 24]. The drying temperatures of flue gas dryers are clearly higher than 100^{o}C (see Table 1). Exhaust gases or unclean condensates must be treated after the dryer if they contain high concentration of emissions. Treatment increases drying costs. The need for the treatment of the exhaust gas depends on current regulations.

After the oil crisis, energy prices fell and dryer investments were no longer so profitable. Due to the absence of economic drivers, attention was not paid to the energy efficiency of

CHP plants. It was more important to develop robust combustion technologies. However, economic drivers (e.g. emissions trading, increasing fuel prices) are more favourable for drying at the moment.

1.4 Objective and scope of the thesis

In 1998 a research project called "Enhancement of combined heat and power production at integrated pulp and paper mill" began at the Helsinki University of Technology. The main objective of the project was to study how to enhance combined heat and power production (CHP production) by means of drying at integrated pulp and paper mills. Instead of high-temperature flue gas drying, the idea was to use low-temperature secondary heat as drying energy. The following research questions were set in the project:

1. Utilization of secondary heat in drying
2. Methods to improve the energy efficiency of the dryer
3. Methods to decrease the drying gas demand without increasing the drying temperature
4. Evaluation of quality and quantity of emissions released from biofuels with low-temperature drying
5. Integration of a biofuel dryer into the CHP process in an optimal way
6. Operation of biofuel dryer under variable drying conditions (integrated into the CHP plant) and the need of control final fuel moisture content

In general, secondary heat means the heat which is expelled from a process in own streams without the heat of the product [25]. Secondary heat from processes goes into the secondary heat system, which is a water-based heat supply system [25]. Temperatures of secondary heat flows are usually lower than 100^{o}C, which means that air is the only realistic drying gas. Heat from secondary heat flows is transferred to air in indirect heat exchangers. In addition, backpressure and/or extraction steam may be used for the heating of air to get higher drying temperatures, if it is reasonable. A conveyor/fixed bed dryer is particularly recommended for low-temperature drying (see Table 1) [14,18].

Several flow sheets for decreasing the heat consumption in air drying may be devised [26]. In the above-mentioned research project, a multi-stage drying system (MSD) was selected as the primary method of improving the energy efficiency in drying and reducing the drying air demand.

Dr Jukka-Pekka Spets worked as a research scientist in the above-mentioned project, and his doctoral thesis "Enhancement of The Use of Wood Fuels in Heat and Power Production in Integrated Pulp and Paper Mill [27]" was published in 2003. In [27], an in-depth performance analysis of the multi-stage drying system was made from thermodynamic point of view. Qualitative and quantitative analyses of emissions released from biofuels with low temperature drying based on experimental tests have also been presented in [27] .

The drying system considered in this thesis is classified as follows: **Drying medium:** Air **Heat supply into the material:** Convection **Transport mechanism of the material inside the dryer:** Conveyor **Method to improve energy efficiency:** Multi-stage drying. Both secondary heat and steam can be used for heating of drying air.

The main objective of this thesis is to analyze research questions 5 and 6. An optimization model for analyzing the integration of a multi-stage drying system into the CHP-process has been developed. A simulation model based on the optimization model has been created to analyze the operation of the dryer under variable drying conditions. The simulation model can also be applied to model-based control of the dryer. Conclusions on the need to control the final fuel moisture content are drawn by comparing simulation results, which are calculated for dryers with and without control. Both models are applied to a case study and results for the integration and operation of the dryer are presented in Chapters 4.2 and 4.3, respectively. In connection with the simulation model, a simple method for approximately determining the initial fuel moisture content is also presented.

To be able to analyze research questions 5 and 6, the drying phenomenon of wood-based material must be known, and guidelines for dimensioning the chosen drying system must be determined. To study the drying phenomenon a series of experimental drying testes have been conducted in a laboratory. Results of experimental tests are presented in Chapter 2. Guidelines for dimensioning a continuous conveyor dryer in the case of multi-stage drying are presented in chapter 3.

Finally, evaluation methods for energy efficiency are discussed. In Chapter 5, energy efficiency of multi-stage drying is analyzed using two evaluation methods: Specific heat consumption and the irreversibility rate. Theoretical minimums for heat consumptions and irreversibility rates are presented and the main difference between the two evaluation methods is clarified. The energy efficiency of multi-stage drying is also compared to single-stage drying with partial recycling of drying air, which is an alternative flow sheet for improving the energy efficiency in drying.

The main results of this thesis are presented in Chapters 2-5, and they are based on six appendix papers:

I Holmberg H, Ahtila P. Drying Phenomena in a Fixed Bed Under Biofuel Multi Stage Drying. In: Oliveira A, Afonso C, Riffat S, editors. Proceedings of the 1st International Conference of Sustainable Energy Technologies, Porto, Portugal; June 12-14, 2002. p. EES1 6-11.

II Holmberg H, Ahtila P. Comparison of drying costs in biofuel drying between multi-stage and single-stage drying. Biomass&Bioenergy 2004; 26: 515-530.

III Holmberg H, Ahtila P. Optimization of the bark drying process in combined heat and power plant process of pulp and paper mill. In: Odilio AF, Eikevik TM, Strommen I, editors. Proceedings of the 3rd Nordic Drying Conference, Karlstad, Sweden; June 15-17, 2005.

IV Holmberg H, Ahtila P. Adjusting of temperature levels in multi stage drying system by means of outlet air measurements. In :Oliveira A, Afonso C, Riffat S, editors.

Proceedings of the 1st International Conference of Sustainable Energy Technologies, Porto, Portugal; 12-14 June, 2002. p. EES2 1-5.

V Holmberg H, Ahtila P. Simulation model for the model-based control of a biofuel dryer at an industrial combined heat and power plant. Drying Technology 2006; 24:1547-1557.

VI Holmberg H, Ahtila P. Evaluation of energy efficiency in biofuel drying by means of energy and exergy analyses. Applied Thermal Engineering 2005; 25: 3115-3128.

1.5 Literature review

The aim of this chapter is to give a brief review of the most relevant research projects related to biofuel drying under the topic of this thesis. Because there are several pulp and paper mills in Finland and Sweden, and both countries use a lot of biofuels, the most relevant studies are done either in Finland or in Sweden. Especially, the Swedish foundation Värmeforsk has funded a lot of studies dealing with the forest industry and biofuels.

References [28-30] discuss utilization of secondary heat at integrated pulp and paper mills based on five case studies. The papers present secondary heat balances of the mills, proposals for more efficient use of secondary heat, and economic analyses of secondary heat investments. There are substantial amounts of unused secondary heat available in all mills. For example, in the mill analyzed in [29] there are hot waters at the temperature of 80^{o}C and temperature range of 57-70^{o}C available 44kg/s and 171kg/s, respectively. However, all available secondary heat flows cannot be used for all purposes. Utilization is always mill-specific, depending on the temperature of the secondary heat flows, mill layout, process concepts, duration curves etc. Biofuel drying is not mentioned as an option for the utilization of secondary heat in [28-30].

In [18], the technical and economic aspects of utilizing secondary heat for biofuel drying has been studied in cases representing both saw mills and pulp and paper mills. A survey of available secondary heat sources at pulp and paper mills and has been made, and commercial drying technologies suitable for biofuel drying have been discussed. The profitability of drying has also been evaluated for single-stage drying. The study resulted in the following main conclusions: 1) There are large sources of secondary heat available for drying purposes at pulp and paper mills. 2) A continuous fixed bed dryer seems to be the most suitable drying technology for biofuel drying at low drying temperatures. 3) Payback periods for the dryer investment are between 2.5-3.4 years depending on the selected initial values. The results presented in [18] support the results of our study. However, several issues discussed in this thesis (e.g. multi-stage drying, design guidelines for the fixed bed dryer, and integration of the dryer into the CHP process) are beyond the scope of [18].

In [17] and [19], operational experiences of existing biofuel dryers have been studied. In most cases, drying is related to pellet manufacturing processes. Both steam and flue gas/air dryers have been analyzed in [17] and [19]. The studies pose questions about raw material, feeding and discharge systems of dried goods, gas and dust handling, fires and control system. According to the results, flue gas/air dryers are more robust than steam dryers. Corrosion and erosion problems and leakage from the feeding and discharge systems have been the most typical problems of steam dryers. From the viewpoint of this thesis, the most relevant information relating to operation experiences is that an air dryer operating at a drying temperature of 100-120^{o}C has had no environmental problems or fires [17]. The dryer is a continuous fixed bed dryer and is used for bark drying at Rockhammars Bruk in Sweden.

In references [31-33], and also in reference [19], the authors discuss control methods for both flue gas and steam dryers used in the pellet manufacturing. In most cases, moisture control of the outgoing material is based on measurement of the exhaust gas temperature. Reference [33] presents a control method based on the connection between the humidity in the flue gas and the moisture in the biofuel. According to references, control methods seem to work quite well. However, the control methods presented for biofuel dryers in [19] and [31-33] are not related to drying in the connection of the CHP plant. They are also based on conventional feedback control, and not on the simulation model as in this book.

Reference [34] contains a literature review of methods for determining moisture content in the biofuel flow. Scanning with Radio Frequent Electromagnetic Waves (RF), Near Infrared Spectroscopy (NIR), Microwaves and Nuclear Magnetic Resonance (NMR) are mentioned in [34] as conceivable methods for determining moisture in the biofuel flow. Because of several reasons mentioned in [34] it is hard to reliably compare the different methods. According to [34] NIR technology can be regarded as the most promising method for the determination of moisture in the biofuel flow.

In [35], the Technical Research Centre of Finland (VTT) has studied drying of wood-based fuel particles both theoretically and experimentally. The drying models presented are based on the solution of the conservation equations for mass and energy. Theoretical drying models for a single particle and for particles in fixed and moving beds are presented in [36, 37]. In both papers, the models are presented as a one-dimensional case for platy, cylindrical, or spherical particles. Experimental tests are carried out by using a thermo balance reactor for drying studies on a single particle. Experimental results are presented in [35-38]. In addition to [35-38], drying of wood-based fuel particles has been studied both theoretical and experimental in [39,40]. Both papers present a two-dimensional drying model for the transport mechanism of a single particle. In [41], a comprehensive heat and mass transfer model is presented for fixed-bed drying of spherical particles. The drying models presented might have been useful methods to calculate drying times in this study, too. However, drying times for the chosen biofuels have been determined experimentally. The decision to opt for experimental tests rather than theoretical models is explained in Chapter 2.3.

2 Drying as a physical phenomenon

2.1 Theory

All biofuels are porous hygroscopic material. Moisture inside the porous material may be in the frozen, liquid or vapor state; in addition, there may be some non-condensable gas present [42]. The transfer of moisture inside the material may occur by several mechanisms, such as diffusion in continuous homogenous solids, capillary flow in granular and porous solids, flow due to shrinkage and pressure gradients, and flow caused by sequential of vaporization and condensation [43]. It is common that different transport mechanisms predominate at different times in a drying cycle. The factors governing heat and mass transfer determine the drying rate [43].

Drying is usually divided into three successive periods: 1) the warming-up period, 2) the constant rate period, and 3) the falling rate period. The warming-up period is normally short compared with the constant and falling rate periods. In the constant rate period, the surface contains free moisture and vaporization takes place from the surface. Toward the end of the constant rate period, moisture has to be transported from inside of the solid to the surface by capillary forces [26]. In the constant drying period, drying is externally controlled by heat and mass transfer resistance on the gas side only. When the average moisture content reaches the critical moisture content, the falling rate period begins. In the falling rate period, drying is controlled by diffusion of moisture from the inside to the surface and then mass transfer from the surface to the drying medium [26]. Drying is said to be internally controlled.

In the constant drying period, heat and mass transfers are in equilibrium at the particle surface. If heat is supplied by convection from hot air, the drying rate may be calculated as follows:

$$\dot{m}_e = \frac{\alpha(t_a - t_s)}{l_v(t_s)} A_e \,, \qquad (1)$$

where α is the heat transfer coefficient, t_a the air temperature, t_s the surface temperature of the particle, A_e the evaporation surface and $l_v(t_s)$ vaporization heat of the water at the surface temperature. In the constant drying period, the surface temperature is the same as the wet bulb temperature. If there are several particles in a fixed-bed, and all particles have reached the local wet bulb temperature, the temperature difference in (1) can be replaced as a logarithmic mean value temperature:

$$\Delta t_{ln} = \frac{(t_{a,in} - t_{s1}) - (t_{a,out} - t_{s2})}{In \dfrac{(t_{,ain} - t_{s1})}{(t_{a,out} - t_{s2})}} \,, \qquad (2)$$

where $t_{a,in}$ is the inlet air temperature, $t_{a,out}$ the outlet air temperature, t_{s1} the surface temperature in the bottom part of the bed, and t_{s2} the surface temperature in the upper part of the bed.

Based on the analogy between heat and mass transfer and the diffusion theory in the boundary layer, the theoretical wet bulb temperature may be derived from Equation (1). The derivation is presented, for example, in [44], and the wet bulb temperature may be calculated from the following equation [44]:

$$t_a - t_{wb} = \frac{M_v}{\rho c_p}\frac{p_o}{RT} Le^{1-n} l_v(t_{wb}) \ln \frac{p_o - p_v}{p_o - p_v^{'}(t_{wb})} \quad , \qquad (3a)$$

where

$$\rho c_p = \rho_{da} c_{pda} + \rho_v c_{pv} \quad , \qquad (3b)$$

$$Le = \frac{\rho c_p}{\lambda} D_{AB} \quad , \qquad (3c)$$

where ρ is the density, c_p the specific heat capacity, Le the Lewis number, D_{AB} the diffusion coefficient of the vapour-air mixture and λ the heat conductivity of the vapour-air mixture. The vapour pressure p_v' as a function of temperature may be expressed using an empirical correlation. Reference [44] gives the following correlation:

$$p_v^{'} = 10^5 e^{\frac{11.78(t_{wb}-99.64)}{t_{wb}+230}} \quad [\mathrm{Pa}] \quad , \qquad (4)$$

The vaporization heat of water as a function of the vaporization temperature may be given approximately as follows [43]:

$$l_v(t_{wb}) = 2501 - 2.34 t_{wb} \qquad [\mathrm{kJ/kg}] \qquad (5)$$

Substituting (4) and (5) in (3a), the wet bulb temperature is the only unknown parameter, and can be calculated from (3a).

2.2 Experimental tests

A series of experimental test was conducted. The tests had the following objectives:

- To determine experimentally the heat transfer coefficient in a fixed bed for wood particles
- To evaluate the applicability of the constant drying model for determining the drying rate in a fixed-bed drying

The tests were conducted using the test rig shown in Figure 3. To calculate the heat transfer coefficient from Equations (1) and (2), the following parameters must be measured: Absolute air moisture contents and temperatures before and after the bed, surface temperatures of particles in the bottom and upper part of the bed, and the evaporation surface. All particles used in the tests were platy spruce particles of regular shape, and the evaporation surface of the bed was determined by multiplying the surface of a single particle by the number of particles in the bed. Other necessary parameters were measured using thermocouples and humidity transmitters (see a more detailed description of the experimental arrangement in Appendix paper I). The measured temperatures and moisture contents used in Equations (1) and (2) represented averages over the period during which the entire bed was in the constant drying period. In all cases, the constant drying period was relatively short, lasting only some ten seconds (see Appendix paper I).

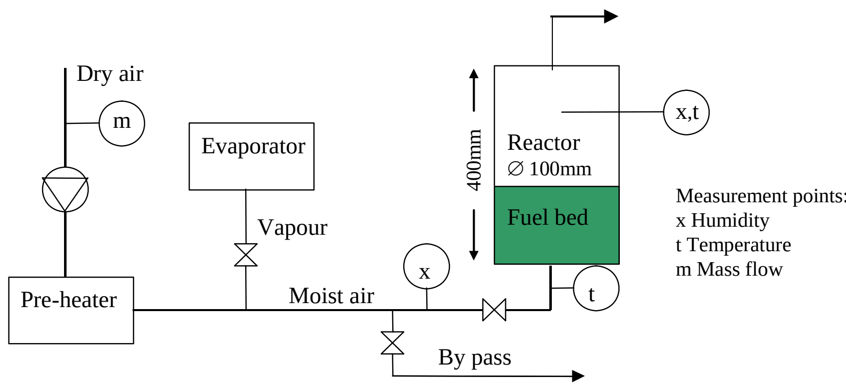


Figure 3. Drying test rig

The particle sizes used in the tests are expressed as effective diameters. The effective diameter is defined as follows [36]:

$$D_{eff} = \frac{6V}{A_e} \quad , \tag{6}$$

where A_e is the total evaporation surface of the particle, and V the volume of the particle.

Heat transfer coefficients were determined for four different particle sizes. Effective diameters and actual dimensions (thickness*length*width) of the used particles were 5mm (2.5*10*10mm), 10mm (20*20*5mm), 15mm (7.5*30*30mm), and 20mm (10*40*40mm). Spruce particles were used because most of the raw wood consumed by the forest industry in Finland is softwood.

Two different inlet air temperatures, 70 and 120°C were used. With the inlet temperature of 70°C, the inlet air was dry. With the inlet temperature of 120°C the inlet air moisture content was around 20 $\pm$ 4g/kg$_{d.a.}$ The air was moisturized with an inlet temperature of 120°C to simulate drying conditions corresponding to the multi-stage drying system (a description of the system is presented in Chapter 3.1)

The air velocities used in the tests were 0.3 m/s, 0.45 m/s (only with the inlet temperature of 70°C), 0.6 m/s, 0.9 m/s and 1.2 m/s. All velocities were calculated per free sectional area of the grate. Because the air density falls as the air temperature rises, the mass flows were smaller with the inlet temperature of 120°C.

The experimentally determined heat transfer coefficients for inlet temperatures of 70°C and 120°C are shown in Figure 4. The measured heat transfer coefficients are smaller for the inlet temperature of 120°C. For both inlet air temperatures, the air velocity before the bed has been the same but the average velocities over the bed have not been the same. As a result of a more drastic temperature drop over the bed, the average velocity has been smaller for the inlet temperature of 120°C. The lower average velocity is probably the main reason, why heat transfer coefficients differ depending on the inlet air temperature.

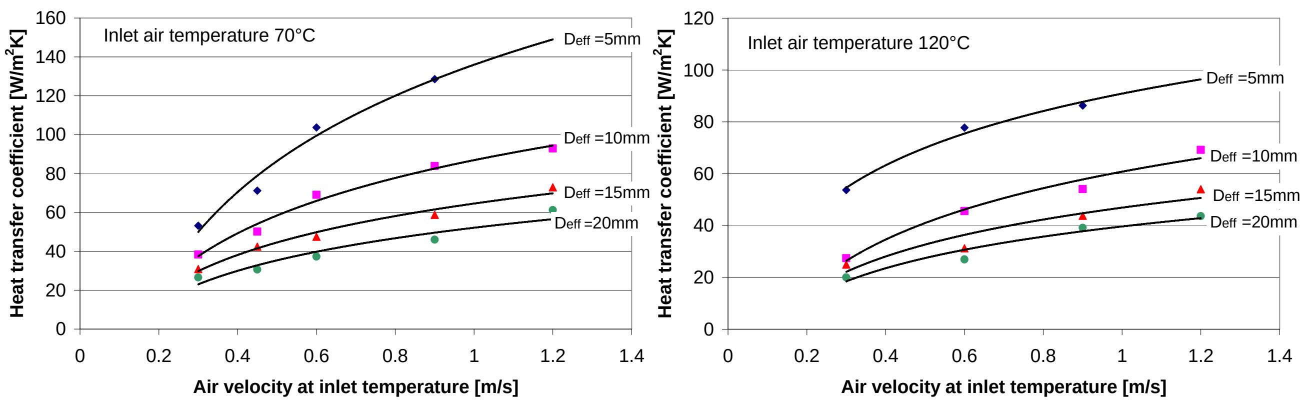


Figure 4. Experimentally determined heat transfer coefficients for effective diameters of 5, 10, 15 and 20mm at inlet temperatures of 70°C and 120°C, D_{eff} = effective diameter (see Equation 6).

To evaluate the validity of the measured heat transfer coefficients they have been compared to heat transfer coefficients calculated using two different Nusselt number correlations. The first correlation is the Ranz-Marshall correlation (Nu_1), and the second one is the Malling and Thodos correlation (Nu_2). The correlations are in the following form [36,45]:

$$Nu_1 = 2\varepsilon + 0.69(Re/\varepsilon)^{1/2}(Pr)^{1/3} \quad (7a)$$

$$Nu_2 = 0.539(Re)^{0.563}(Pr)^{1/3}/\varepsilon^{1.19} \quad (7b)$$

Comparison to these two Nusselt number correlations has been done for cases where the inlet air temperature is 70°C. Reynolds and Prandtl numbers have been determined at the surface temperature of the particle. The surface temperature used represents the average of the measured surface temperatures in the bottom and upper part of the bed. Depending on the measurement, the surface temperatures used varied between 21.5-27.5°C. For small particles surface temperatures are higher. The properties (e.g. viscosity, conductivity) of moist air are taken from [42]. The volume fraction of air ε is 0.57 in the Nusselt number correlations, and has been determined experimentally by filling a measure of volume with particles. The results of the comparison are shown in Figure 5. Because the number of pages was limited in Appendix paper I, Figure 5 was excluded in this paper.

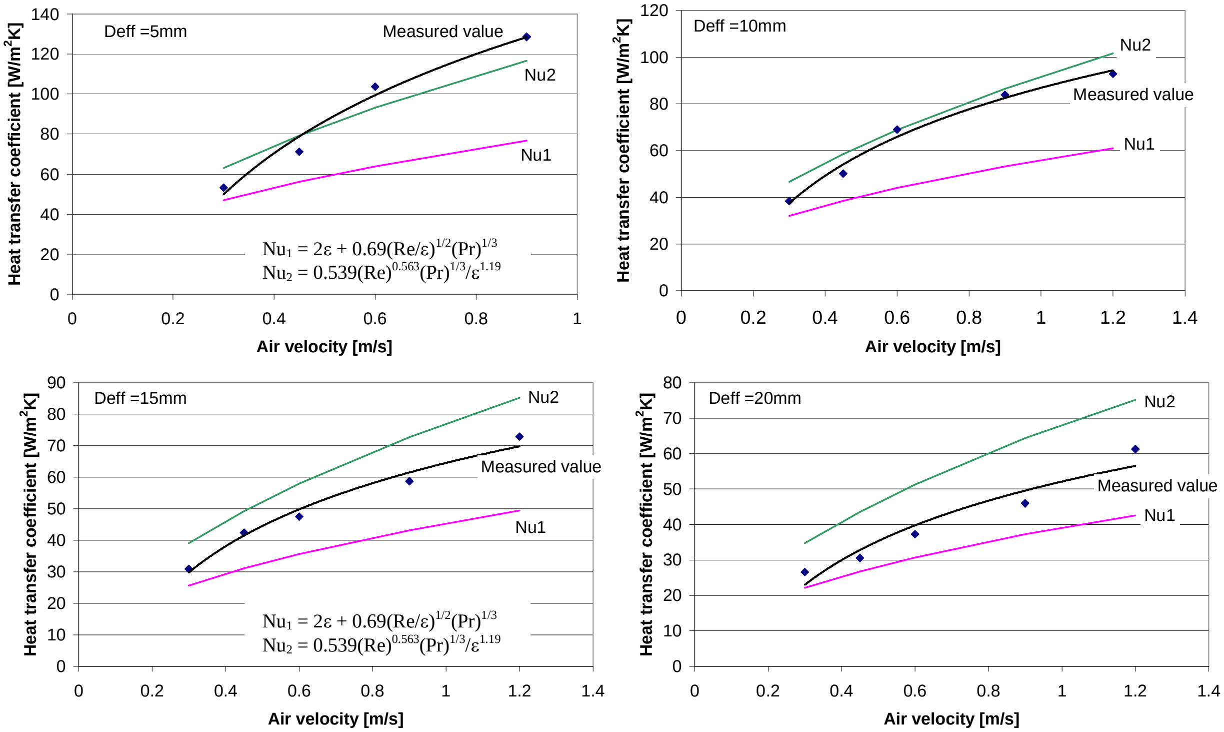


Figure 5. Comparison of measured heat transfer coefficients at an inlet air temperature of 70^oC with coefficients calculated using Ranz-Marshall correlation (Nu_1) and Malling and Thodos correlation (Nu_2), D_{eff} = effective diameter (see Equation 6).

Figure 5 shows that experimentally measured heat transfer coefficients correspond to calculated heat transfer coefficients reasonably well. Especially for small particles, the Malling and Thodos correlation is good. On the basis of Figure 5 it is reasonable to assume that the validity of the measured heat transfer coefficients is also good for the inlet temperature of 120^oC.

To evaluate the applicability of the constant drying model for determining the drying rate, the constant drying model was used to calculate the decrease in the moisture content of the particles. The bed was divided into ten layers, and the evaporation surface was assumed to be equal in each layer. In each layer, the drying conditions were assumed to be constant and the air temperature and moisture content of the upper layer was calculated by applying the energy and mass balance of adiabatic moisturizing. Experimentally determined heat transfer coefficients were used in the model. The calculation model is explained in more detail in Appendix paper I.

The reduction in the moisture content calculated by the constant drying model was then compared with thee measured value. The effect of air velocity and particle size on the critical sample moisture content was evaluated. In this connection, the critical sample moisture content refers to average sample moisture content below which the constant drying model is no longer accurate. According to the results shown in Appendix paper I, the critical sample moisture content seems to be slightly lower for small particles and air velocities. However, the influences are so small that they can be ignored. Figure 6a shows for all particle sizes a comparison between actual sample moisture contents and theoretical moisture contents calculated using the constant drying model. Each curve in Figure 6a represents already the average of the measurements where air velocity has been a variable. Figure 6b shows the average of the curves shown in Figure 6a, and this curve gives us an estimation of the critical sample moisture content.

Figure 6a and b show that the critical sample moisture content is in the order of 1.0 kg/kg_{dm}, and the constant drying model is no longer accurate, since the sample moisture content has decreased below 0.8-0.9kg/kg_{dm}.

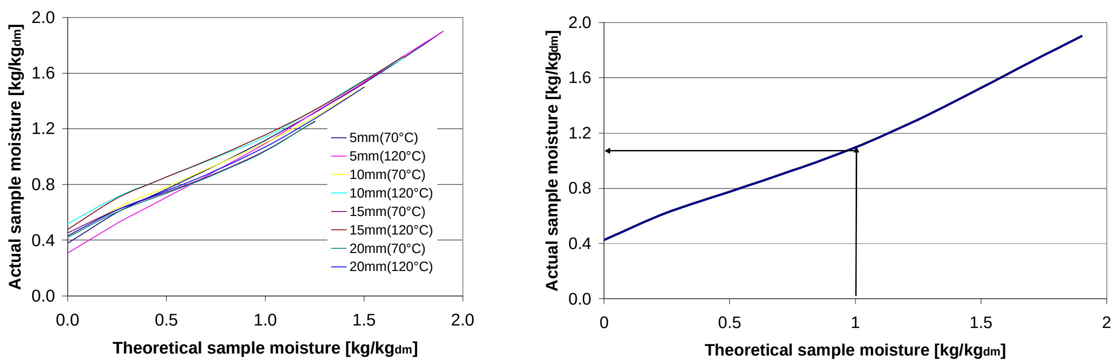


Figure 6. a) The effect of particle size and temperature on the accuracy of the constant drying model. b) The average accuracy of the constant drying model.

2.3 Conclusions on experimental tests

The following conclusions can be drawn from experimental tests

- The stage during which the entire bed is in the constant drying period is short compared with the total drying time.
- Experimentally determined heat transfer coefficients behave logically as a function of particle size and air velocity.
- The Nusselt number correlation of Malling and Thodos determines the heat transfer coefficient with good accuracy for particle sizes of 5 and 10mm.
- The accuracy of the Ranz-Marshall Nusselt number correlation improves as the particle size increases.

- The critical moisture content in a fixed bed is in the order of 1.0kg/kg$_{dm}$ for the spruce particles used in the tests.
- The constant drying model is not accurate enough below the critical moisture, and it is necessary to use diffusion-based drying models to determine the drying time theoretically in a fixed bed, even if the final fuel moisture content is clearly higher than the fiber saturation point.

Even if the accuracy of the constant drying model was better for average moisture contents clearly below 0.8-1.0kg/kg$_{dm}$, the application of the model to actual biofuels would still require further development. The particle size distribution is extremely wide in an actual biofuel flow (see. Figure 2), and it would be necessary to determine the “correct” particle size from the viewpoint of the model. Especially in the case of bark, it is hard to say what the particle size of the fuel is. The use of diffusion-based drying models would require material properties such as diffusivities and conductivities to be determined in addition to the “correct” particle size. In stead of further model development, it was decided that experimentally determined drying curves would be used to determine the drying times.

3 Design of the drying system

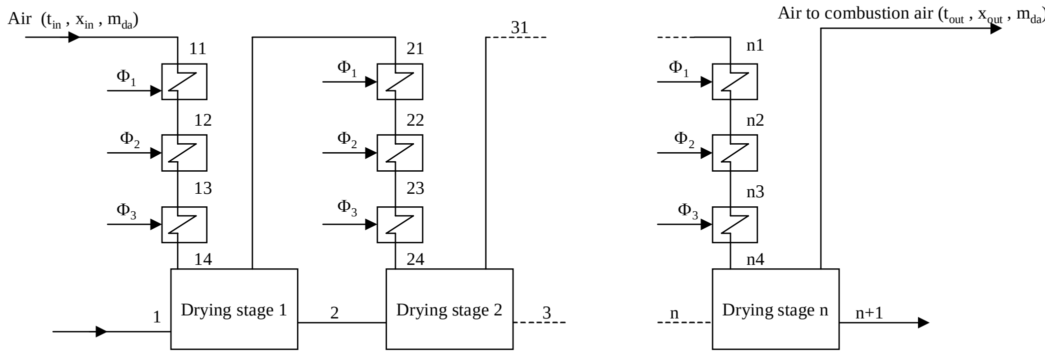
3.1 Multi-stage drying system

The flow sheet of the multi-stage drying system used in the model development is shown in Figure 7. Figure 7b illustrates the states of the drying air on an enthalpy-humidity chart diagram in multi-stage drying. A drying stage consists of the heating and drying period.

The addition of drying stages reduces heat consumption in drying, because the air temperature before heating increases constantly as the number of drying stages increases (see Figure 7b) [46]. Adding the drying stages also decreases the air demand in drying if all the drying stages operate at the same temperature level. The reason is the higher outlet air humidity, which is achieved as a result of several drying stages. To reach the same outlet air humidity in single-stage drying, the drying temperature should be higher than in the case of multi-stage drying (see. Figure 7b) [46]. In Figure 7b, two-stage drying gives the same change of air moisture content as the single-stage drying, but air temperatures are lower in two-stage drying. Although the air demand decreases, the dryer dimensions do not become smaller because the drying air must pass through several drying stages.

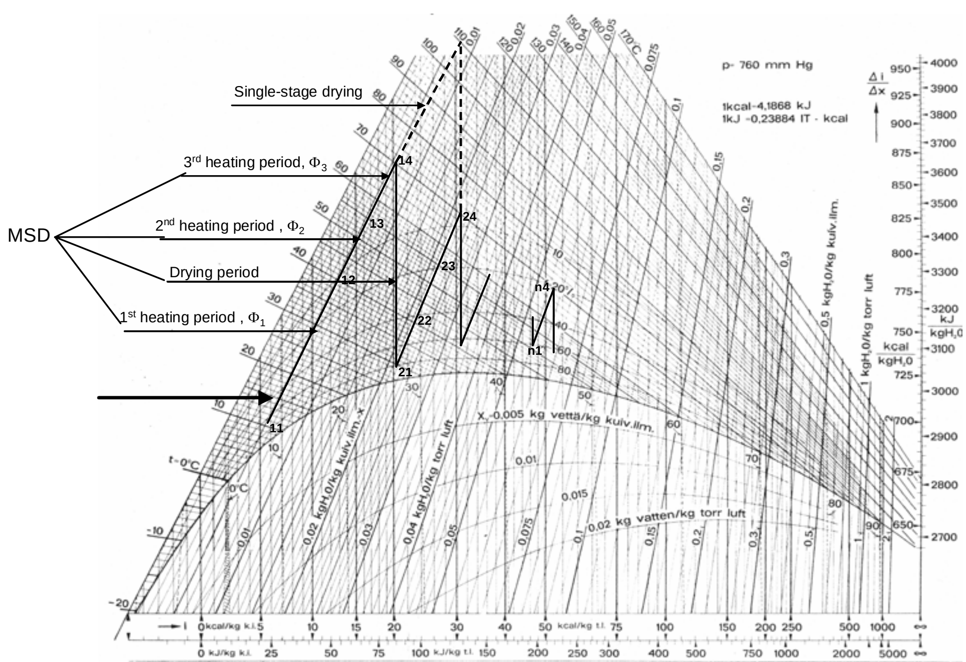
Heating the drying air successively in three heat exchangers does not reduce the heat consumption but decreases the consumption of steam. An optimal combination of heat sources in each drying stage depends on the initial values and is calculated by the model. Thus, there does not have to be three heat exchangers in each drying stage.

a)



Φ_1 Secondary heat , typical temperature range 50-90 °C Φ_3 Back pressure steam , typical temperature range 130-145 °C
Φ_3 Extraction steam , typical temperature range 180 - 190 °C

b)



Heat sources
Φ_1 Secondary heat
Φ_2 Backpressure steam
Φ_3 Extraction steam

Translations:
kuiv. ilm.= *dry air*
torr luft = *dry air*
kg vettä/ kg kuiv.ilm. =
kg water / kg dry air
kg vatten / kg toff luft =
kg water / kg dry air

NOTE!
Due to limited temperature scale of the enthalpy humidity chart, drying temperatures in MSD on are lower than actual ones after the heating with secondary heat, backpressure steam and extraction steam.

Figure 7. Multi-stage drying a) flow sheet b) states of the drying air on an enthalpy-humidity chart diagram.

3.2 Determination of dryer dimensions

Dryer dimensions are determined for a continuous conveyor dryer in which air moves through a perforated conveyor and a fixed fuel bed (see. Fig. 8). Functionally, the dryer corresponds to a cross-flow concept. The drying time from the given initial moisture content to the desired final moisture content is determined experimentally based on the conclusions presented in Chapter 2. In addition to single-stage drying, a concept for determining dryer dimensions in multi-stage drying is presented.

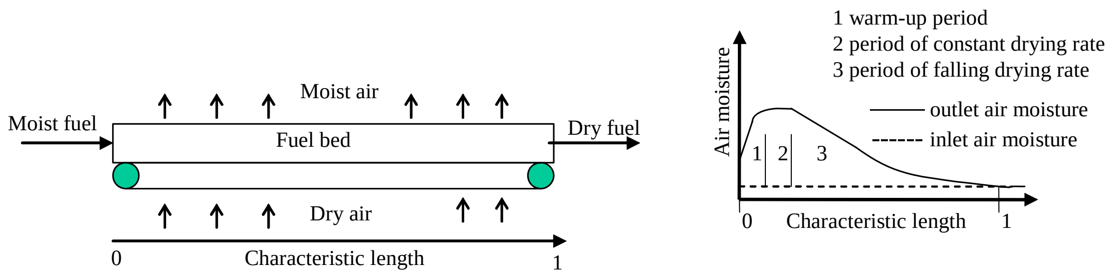


Figure 8. A continuous cross-flow dryer and the change of outlet air moisture content during drying

An equation for determining the mass flow of dry air in continuous cross-flow drying is derived in Appendix paper II, and the final equation takes the following form:

$$\dot{m}_{da} = \frac{v_a \rho_{da} \dot{M}_{dm}}{Z(1-\varepsilon)\rho_{dm}} \tau_u \,, \tag{8}$$

where v_a is the air velocity per free sectional area of the dryer, ρ_{da} the density of dry air, $\dot{M}_{dm}$ the dry mass flow of fuel, Z the bed height, ε the volume fraction of air in the fuel bed, ρ_{dm} the density of dry fuel, and τ_u the drying time from the given initial moisture content to the desired final moisture content.

Experimentally, the drying time can be determined directly or indirectly. In this study, the drying time has been determined indirectly by measuring the change of air moisture content over the fuel bed situated in a test rig (see Figure 3). The decrease in fuel moisture content is calculated as follows:

$$u = u_{in} - \frac{v_a \rho_{da}}{Z\rho_{dm}} \sum_{i=1}^{n} (x_{out} - x_{in})_i \Delta\tau_i \quad , \qquad \tau_u = \sum_{i=1}^{n} \Delta\tau_i \,, \tag{9}$$

where x_{out} is the outlet air moisture content, x_{in} the inlet air moisture, n the number of time intervals, Δτ the length of time interval, and τ_u the total drying time. The length of the time interval was 5 seconds in the experimental tests. Equation (9) represents a drying curve in a mathematical form and the measured curve is analogous to continuous cross-flow drying.

In Equation (8), the air velocity and bed height are known parameters, and the drying curves are measured for the chosen air velocity and bed height. One of the main questions is how to choose these parameters? In the case of air velocity the answer is simple. The air velocity must be as high as possible, because the cross-sectional area of the dryer is inversely proportional to the air velocity. In fixed bed drying, the air velocity must be slightly lower than the minimum fluidized velocity. The minimum fluidized velocity depends on the particle size, particle density, and volume fraction of air in the bed [46]. For example, Reference [47] presents an equation for the theoretical determination of the minimum fluidized velocity. In practice, the correct air velocity per free sectional area must be determined experimentally for biofuels. Velocities between c. 0.4-0.6m/s are reported for fixed bed dryers in Ref. [14]. In this study, the air velocity at the inlet temperature of drying air is c. 0.6-0.65m/s in the determination of the drying curves (see Figure 10).

Equation (8) shows that the air mass flow is inversely proportional to the bed height. On the other hand, the drying time becomes longer as the bed height increases. To evaluate the effect of bed height on air mass flow, the behavior of the ratio between the drying time and the bed height (τ_u/Z) was experimentally studied. Regularly shaped wood particles were dried in a fixed bed reactor (see Figure 3) by changing the bed height and keeping the other drying parameters constant. All the particles were ideal spruce particles of the same size and dimensions (2x20x5mm). Two different air temperatures, 70^{o}C and 120^{o}C, were used. Figure 9 shows the results of the tests.

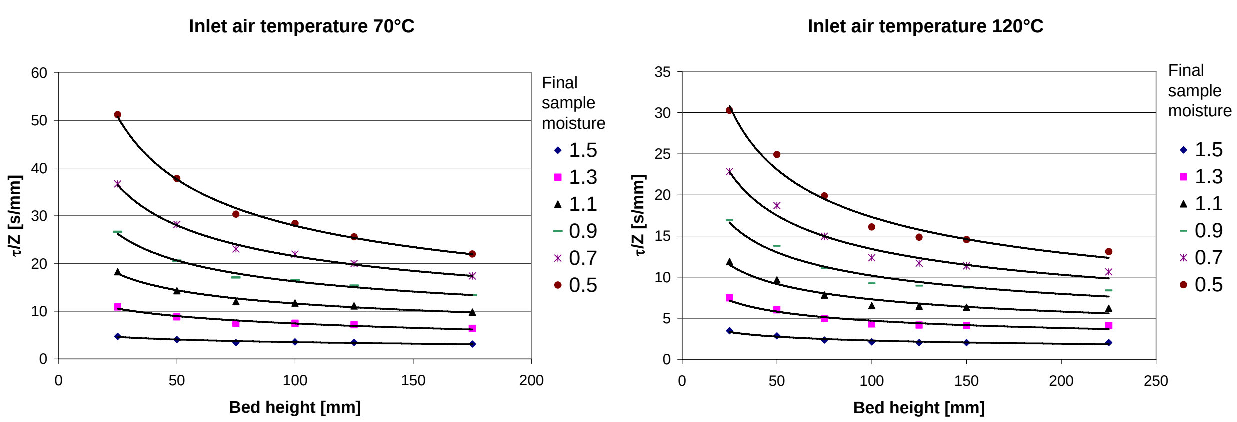


Figure 9. The ratio between drying time and bed height as a function of bed height for inlet temperatures 70^{o}C and 120^{o}C (air velocity 0.6m/s, inlet air moisture 3±0.5g/kg$_{da}$, initial sample moisture 1.7kg/kg$_{dm}$).

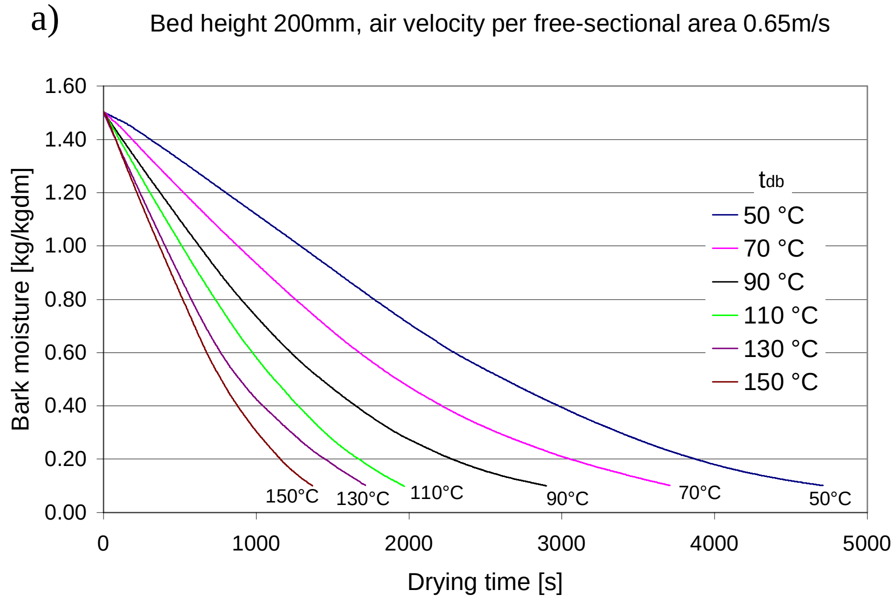
Despite the final moisture content of the sample, the ratio τ_u/Z seems to decrease constantly as the bed height increases. However, the derivative of the ratio is clearly smaller when the bed heights are over 80-100 mm and 100-120 mm for inlet temperatures of 70 and 120^oC, respectively. Both bed thicknesses approximately represent bed heights when the outlet air is fully saturated at the beginning of drying (see Figure 8 in Appendix paper II). On the basis of these measurements, the bed must be at least so high that the drying air reaches its saturation point at the beginning of the drying. Increasing the bed height still decreases the dryer size but the influence on dimensions is not as significant as for thin beds. Bed heights between 0.2-0.8m are reported for conveyor dryers in Ref. [14]. To achieve a uniform air distribution over the fuel bed it may be necessary to use higher bed heights close to reported ones in actual dryers.

Figure 10 shows experimentally determined drying curves when bark (see Figure 2a) has been dried in a fixed bed dryer. Figure 10a shows drying curves as a function of dry bulb temperature, and Figure 10b shows the same curves as a function of the difference between dry and wet bulb temperatures. Equations (3)-(5) show that the temperature difference depends on both dry bulb temperature and inlet air moisture content. To take into account the effect of air moisture content on drying time, the difference between dry and wet bulb temperatures is used in calculation models instead of merely the dry bulb temperature. Drying time τ_u for a certain change of fuel moisture content is expressed using a correlation based on experimental determined drying curves. The correlation is defined as follows:

$$\frac{\tau_u}{\tau_{u,\max}} = \ln\left[\frac{t_{db} - t_{wb}}{(t_{db} - t_{wb})_{\max}}\right]^a + b \ , \qquad (10)$$

where t_{db} is the dry bulb temperature, t_{wb} the wet bulb temperature and $\tau_{u,max}$ the drying time for the smallest temperature difference (t_{db}-t_{wb}) when fuel is dried from a given initial moisture to a desired final moisture. The term $(t_{db}-t_{wb})_{max}$ is the greatest temperature difference used in the experiments. Coefficients a and b are calculated from experimental drying curves based on the minimization of the partial least square, which can be easily done by the computer.

a)



b)

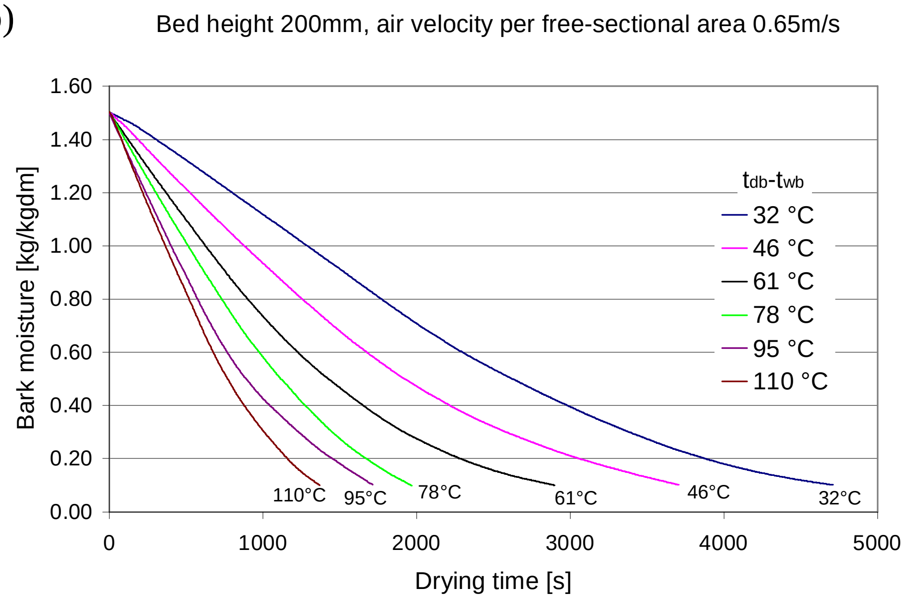


Figure 10. Experimentally determined drying curves a) as a function of dry bulb temperature b) as a function of the difference between dry and wet bulb temperatures.

Guidelines for determining the dryer size in the case of single-stage drying are now presented. In multi-stage drying, the final fuel moisture content u_{out} is achieved after the last drying stage, and the fuel moisture content after the previous stages is somewhere between u_{in}-$u_{out.}$ To use the Equation (8) in multi-stage-drying, the fuel flow $\dot{M}_{dm}$ between each drying stage is shared as shown in Figure 11. The change in fuel moisture content for the shared flow in each drying stage is u_{in}-$u_{out,}$ but the fuel flow is smaller. The moisture balance over the drying chamber is:

$$\dot{m}_{da}(x_{out} - x_{in}) = \dot{M}_{dm}(u_{in} - u_{out}) \tag{11}$$

Substituting Equation (8) in (11) the air moisture content after the drying chamber can be calculated. The inlet air moisture content before the first drying stage is a known initial value and the calculated outlet air moisture is the new inlet moisture content for the second drying stage. Knowing the inlet air temperature and other process parameters in the second stage, the wet bulb temperature and drying time can be determined from Equations (3)-(5) and (10). Then the same actions as in the second stage are repeated for the remaining stages. The air mass flow through the multi-stage drying system is constant and the values for parameters y_1-y_n are calculated as follows:

$$y_1 \frac{v_{a1}\rho_{da1}\dot{M}_{dm}}{Z_1(1-\varepsilon_1)\rho_{dm}}\tau_{u1} = y_2 \frac{v_{a2}\rho_{da2}\dot{M}_{dm}}{Z_2(1-\varepsilon_2)\rho_{dm}}\tau_{u2} \tag{12a}$$

$$\vdots$$

$$y_{n-1} \frac{v_{an-1}\rho_{dan-1}\dot{M}_{dm}}{Z_1(1-\varepsilon_{n-1})\rho_{dm}}\tau_{un-1} = y_n \frac{v_{an}\rho_{dan}\dot{M}_{dm}}{Z_n(1-\varepsilon_n)\rho_{dm}}\tau_{un} \tag{12b}$$

$$y_1+y_2+\ldots y_n = 1 \tag{12c}$$

It is important to note that sharing the fuel flow is only a computational action, and there is no reason to share the fuel flow in a real dryer. Since the values for parameters y_1-y_n are known, approximative fuel moisture contents between drying chambers may be calculated as follows:

$$u_{1out} = u_{1in}\text{-}y_1(u_{in}\text{-}u_{out}) \tag{13a}$$

$$u_{2out} = u_{2in}\text{-}y_2(u_{in}\text{-}u_{out}) \;,\; u_{2in} = u_{1out} \tag{13b}$$

$$\vdots$$

$$u_{n,out} = u_{n,,in}y_n(u_{in}\text{-}u_{out}), \quad u_{n,in} = u_{n-1,out} \tag{13c}$$

Basically, the fuel flow can be shared in a real dryer, too. However, sharing the fuel flow in practice may be difficult to carry out.

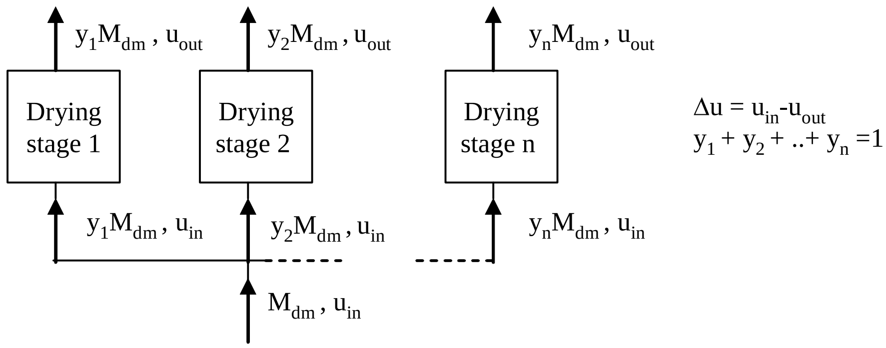


Figure 11. Sharing the fuel flow between drying stages in n-stage drying.

3.3 Conclusions on the dryer design

The following conclusions can be drawn concerning the design of the dryer:

- In the design of the conveyor/fixed bed dryer, the air velocity must be slightly lower than the minimum fluidized velocity and the bed thickness must be at least so high that the drying air reaches its saturation point at the beginning of the drying.
- Increasing the bed height still decreases the dryer size but the influence on dimensions is not as significant as for thin beds.

Calculation equations for determining the dryer dimensions in single- and multi-stage drying are also presented. The determination of dryer dimensions is based on the use of the wet bulb temperature and fuel flow sharing.

4 Integration and operation of the biofuel dryer in an industrial CHP plant

4.1 Industrial CHP plant

Figure 12 shows the energy and material flows in and out of the control region of a typical CHP plant with a fluidized bed boiler in an integrated pulp and paper mill. In addition to the fluidized bed boiler, there are other boilers (e.g. a black liquor boiler, a gas turbine + heat recovery boiler, and an oil boiler for peak load situations) at the mill, too. The structure of the entire energy generation system depends on the products the mill produces. For example, there is always a black liquor boiler if the mill produces chemical pulp. If the mill produces mechanical pulp, a gas turbine + heat recovery boiler may be found at the mill. To discuss the questions as they are raised in this thesis, it is enough that we focus only on the CHP plant shown in Figure 12.

The total efficiency of the CHP plant is defined as follows:

$$\eta = \frac{P + \Phi_h}{Q_f} \quad , \tag{14}$$

where P is the power generation, Φ_h the process heat generated in the form of steam, and Q_f the fuel input into the fluidized bed boiler. The power-to-process heat ratio in CHP production is defined as follows :

$$\alpha = \frac{P}{\Phi_h} \tag{15}$$

The power-to-process heat ratio and efficiency are known parameters for each CHP plant. The fuel consumption can be calculated from equations (14) and (15) as the heat consumption is known.

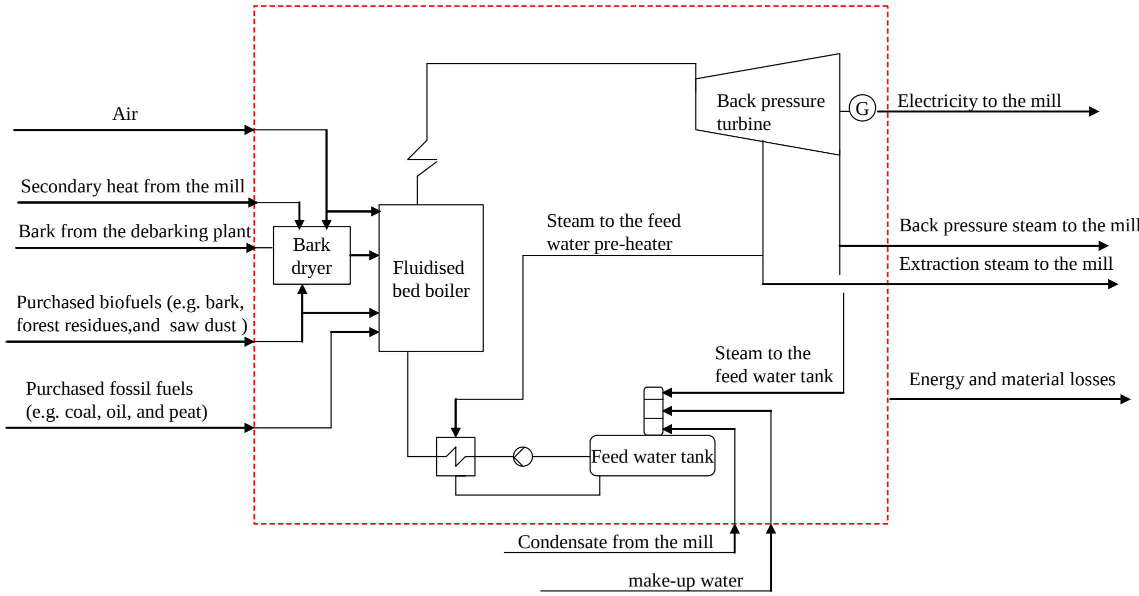


Figure 12. Energy and material flows in and out of a CH -plant at a pulp and paper mill.

The main purpose of the CHP plant is to produce enough steam to cover the heat consumption of the mill at all times. The steam consumption at the mill is not constant but varies constantly depending on the production volume of the mill, the weather and the season. Variation in steam generation means that the fuel input into the boiler varies, too. Usually, fuels are consumed in the following order: first, black liquor (if chemical pulp is produced), then bark from the debarking plant, and finally purchased fuels. Purchased fuels can be both bio and fossil fuels, and the CHP plant can use more than one purchased fuel. However, it is necessary to mention that the fuel consumption does not always follow the foregoing order. For example, the availability of purchased fuels, operation hours of the debarking plant, and problems caused by the high moisture content of the bark may affect the order of fuel consumption.

4.2 Integration of the dryer into the CHP process

4.2.1 Model description

When the biofuel drying system discussed in this thesis is integrated into a CHP process, the following three questions will be raised:

1. What will the optimal final fuel moisture content be?
2. What heat sources will be used for the heating of drying air: only secondary heat, secondary heat + steam or only steam?
3. What will the optimal number of drying stages be?

An optimization/calculation model has been created for analyzing questions 1-3. The model has been developed using the calculation principles presented in chapter 3.2.

The objective function in the model is the net present value (NPV), which takes into account all cash flows over the economic lifetime of the dryer. Taking investment costs and maintenance costs as negative cash flows, the net present value for the dryer becomes [48]:

$$NPV = \sum_{\tau=0}^{\tau=k} \frac{(NI\tau_{op} - C_{maintenance})}{(1+i)^{\tau}} - C_{investment} \quad , \qquad (16)$$

where NI is net income of drying in a time unit, k the total number of years over which cash flows occur, and τ_{op} the annual operating hours of the dryer. Investment costs are expressed as a function of a suitable capacity factor. For a continuous conveyor dryer operating as a cross-flow concept, a suitable capacity factor is the air mass flow or a parameter which can be derived from the air mass flow (e.g. cross-sectional area or volume flow). If the dryer uses secondary heat, backpressure and extraction steam, the net income (NI) of drying in a time unit is calculated as follows:

$$\begin{aligned} NI = \dot{M}_{dm}(u_{in} - u_{out})l_v b_{mf} - \Phi_{sh} b_{sh} - \frac{(\Phi_{bs} + \alpha_{bs}\Phi_{bs})}{\eta} b_{mf} + \alpha_{bs}\Phi_{bs} b_e - \\ \frac{(\Phi_{es} + \alpha_{es}\Phi_{es})}{\eta} b_{mf} + \alpha_{es}\Phi_{es} b_e - \frac{\dot{m}_{da}\Delta p}{\rho_{da}\eta_{fan}} b_e \end{aligned} \quad , (17)$$

where where M_{dm} is the dry mass flow of the fuel through the dryer, u_{in} the initial fuel moisture, u_{out} the final fuel moisture, l_v the vaporization heat of water at a temperature of 25^{o}C, b_{mf} the price of marginal fuel, Φ_{sh} the secondary heat consumption of the dryer, b_{sh} the price of secondary heat, Φ_{bs} the backpressure steam consumption of the dryer, α_{bs} the power-to-heat ratio in backpressure steam generation, η the efficiency of the CHP plant calculated from the fuel input, b_e the price of electricity, Φ_{es} the extraction steam consumption of the dryer, α_{es} the power-to-heat ratio in backpressure steam generation, Δp the pressure drop over the drying system, and η_{fan} the efficiency of the fan.

Equation (17) assumes that the fuel input into the boiler is adjusted on the basis of the heat consumption of the mill and no extra heat is produced for increasing the power generation. This is how CHP plants usually operate at pulp and paper mills. Because drying improves the effective heating value of the biofuel, marginal fuel can be replaced with biofuel. Savings in marginal fuel consumption (the first term in Equation (17)) represent a positive cash flow for a company. If steam is used as drying energy primary heat consumption increases and fuel input into the boiler may increase, too. In this case, earnings stem from the increased power generation. Depending on the initial values used the model calculates whether the earnings stem from the decreased marginal fuel consumption, increased power generation or both. In some cases, a sufficient boiler capacity can be attained by installing a dryer before the existing boiler, and a boiler investment is avoided. In these cases, savings would result from the avoided boiler investment. However, these cases are beyond the scope of this thesis.

Substituting (17) in (16) and expressing investment costs as a function of air mass flow, the objective function is completed. The aim of the model is to determine the final bark moisture content, the combination of heat sources, and the number of drying stages, which maximize the objective function. The entire model with boundary conditions is presented in more detail in Appendix paper III and is not discussed any further here. With regard to the application of the model, can be stated that the model calculates the drying temperatures, heat consumptions, and air mass flow maximizing the NPV in a case where the final fuel moisture content u_{out} and the number of drying stages are given. To determine the final fuel moisture maximizing the NPV, the calculation must be made for different values of u_{out}. On the basis of these values, the NPV can be drawn as a function of $u_{out,}$ and the optimal final fuel moisture content is found graphically (see Figure 13).

4.2.2 Results and discussion

The model has been applied in a case study to analyze questions 1-3 mentioned at the beginning of chapter 4.2.1. In the case study, a virtual dryer is integrated into an existing CHP plant, which is situated at a pulp and paper mill in Finland. The material to be dried is soft wood (pine) bark. Experimentally determined drying curves for the bark are shown in Figure 10, and the same curves have been used in the calculation. The process values of the CHP plant, the temperature level of secondary heat, the initial bark moisture content and the dry mass flow of bark are based on data obtained from the mill where the CHP plant is situated. Prices of marginal fuel and electricity correspond to the price level in 2005. Table 2 shows a summary of all the initial values used in the calculation.

The drying system is assumed to consist of a conveyor, heat exchangers, air ducts, a covering and a fan. Cost functions as a function of air mass flow or cross-sectional area of the dryer (also dependent on air mass flow) are shown in Table 3. Cost functions are based on data obtained from equipment suppliers and reference [48]. To take into account costs such as instrumentation, lagging etc., the sum of the cost functions is multiplied by a factor of 1.6, which is based on information obtained from [48]. The determination of cost functions is explained in more detail in Appendix paper II. Maintenance costs are assumed to be 3 per cent of investment costs.

Table 2. Initial values for calculation cases. Temperatures in parentheses refer to Fig. 7a.

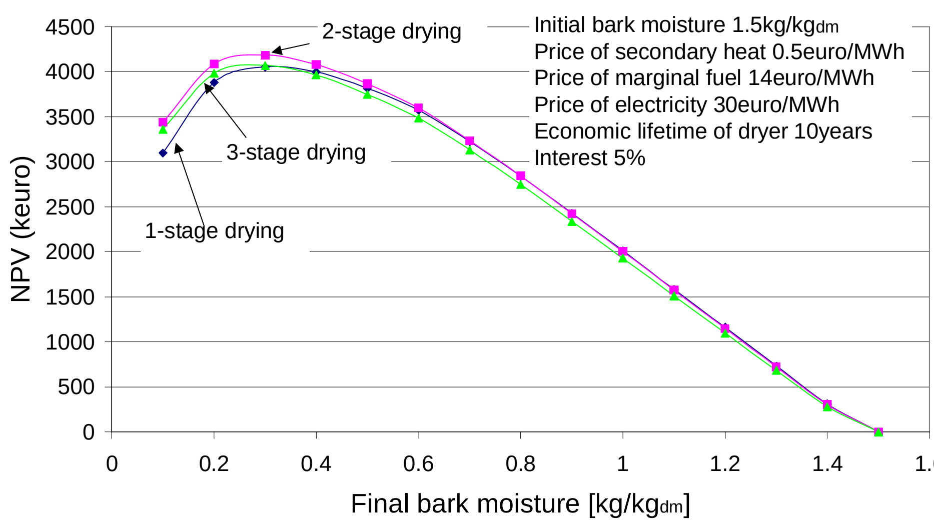
Parameter	Value
Highest drying temperature when secondary heat is used (t_2,t_6)	70°C
Highest drying temperature when backpressure steam is used (t_3, t_7)	120°C
Highest drying temperature when extraction steam is used (t_4 , t_8)	150°C
Maximum number of drying stages	3
Annual operating time of dryer	8400h
Air velocity per free-sectional area of dryer	0.65m/s
Power-to-heat ratio in extraction steam production	0.271
Power-to-heat ratio in backpressure steam production	0.336
Efficiency of CHP-plant (equation (1))	0.87
Pressure drop of each drying stage (estimated)	400Pa
Bed thickness	200mm
Initial bark moisture content	1.5kg/kg_{dm}
Bulk density of dry bark	85kg/m^3
Dry mass flow of bark	4kg/s
Price of electricity	30euro/MWh
Price of secondary heat	0.5euro/MWh
[1]Price of marginal fuel	14euro/MWh
Economic life of dryer	10years
Interest rate	5%
Outdoor temperature (t_1)	5°C
Outdoor humidity (x_1)	0.004kg/kg_{da}

Table 3. Cost functions for main components of the dryer

Equipment	Relationship	Capacity parameter Y	Additional parameter
Conveyor	2700Y	Cross-sectional area	
Air-water heat exchanger	$9\Delta tY^{0.9}$	Air mass flow	Δt is temperature increase in heat exchanger
Air-steam heat exchanger	$18\Delta tY^{0.9}$	Air mass flow	Δt is temperature increase in heat exchanger
Air duct	$3770Y^{0.5}$	Air mass flow	
Fan	$0.9\Delta pY^{0.7}$	Air mass flow	Δp is pressure drop of drying stage
Covering	$1200Y^{0.5}$	Cross-sectional area	

In a basic case, the optimal drying system is determined using the initial values shown in Table 2. Calculation results for 1-, 2- and 3-stage drying systems in a basic case are shown in Figure 13.

The effect of the length of the economic lifetime and prices of marginal fuel, electricity and secondary heat on the optimal drying system is evaluated through the sensibility study. In the sensibility study, one parameter (e.g. the price of marginal fuel) is changed and other parameters have the same value as in Table 2. The following values are used in the sensibility study calculations: length of economic life 1-20 years, the price of marginal fuel 6-20euro/MWh, the price of electricity 20-70euro/MWh and the price of secondary heat 0-3euro/MWh. The price of secondary heat depends on the mill's pricing policy. Some mills include some costs (e.g pumping costs) in the price of secondary heat and others do not price secondary heat at all. The maximum number of drying stages can be 3, and drying systems with more drying stages have not been considered in this study. Figures 14-17 show the results of the sensibility study. The results are calculated in a way similar to that shown in Figure 13 but only the best case is shown in Figures 14-17. If the drying temperature is higher than 70°C, steam must be used for the heating of drying air in addition to secondary heat. However, in some cases it may be more profitable to use only steam, not secondary heat, in the dryer. These cases are marked separately in Figures 14-17.



Calculation results at global optimum

Number of drying stages	Optimal drying temperature [°C]	Optimal final bark moisture [kg/kg$_{dm}$]	Specific heat consumption [kJ/kg$_{H2O}$]	Investment costs [keuro]	Maximum net present value [keuro]
1	70	0.3	5860	4300	4053
2	70	0.3	4360	4520	4185
3	70	0.3	3890	4725	4071

Figure 13. Net present values as a function of final bark moisture contet and calculation results at global optimum maximizing the net present value in a basic case (see initial values from Table 2).

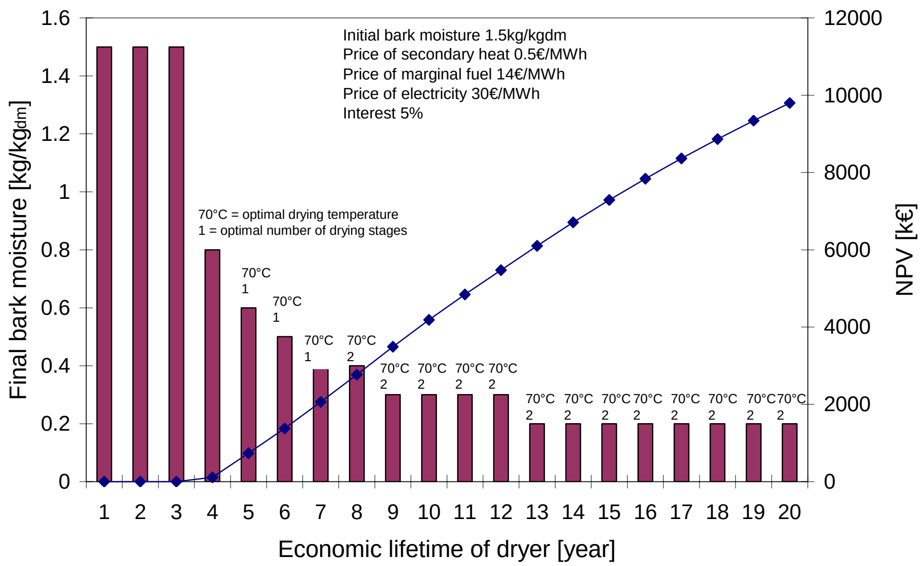


Figure 14. Results of the sensibility study as a function of economic lifetime. Reading example: If the economic lifetime is 5 years (see other initial values from the figure), the figure shows that the optimal final bark moisture is 0.6 (columns), the optimal number of drying stages 1, the optimal drying temperature 70^{o}C and the net present value approximately 500k€ (line).

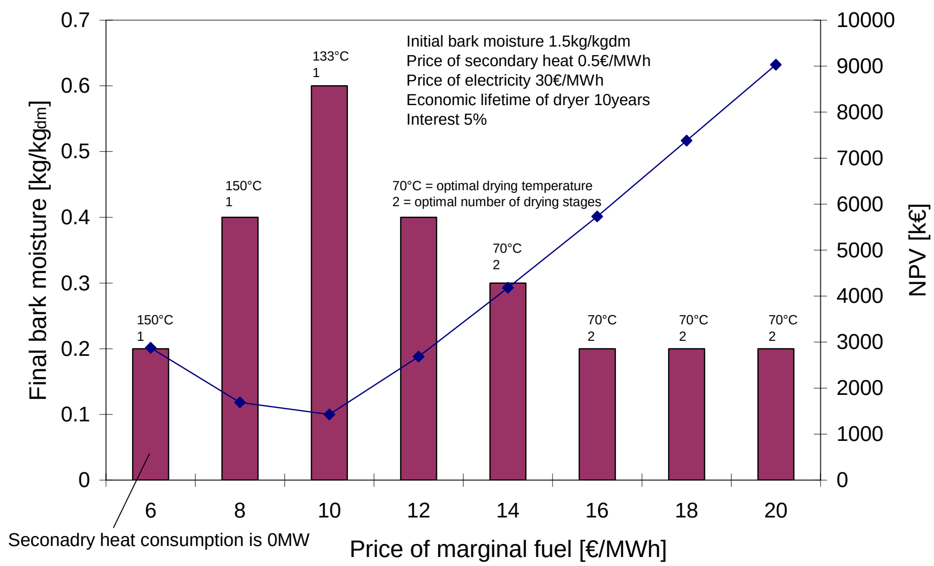


Figure 15. Results of the sensibility study as a function of the marginal fuel price. Reading example: If the marginal fuel price is 16 €/MWh (see other initial values from the figure), the figure shows that the optimal final bark moisture is 0.2 (columns), the optimal number of drying stages 2, the optimal drying temperature 70^{o}C and the net present value approximately 6000k€ (line).

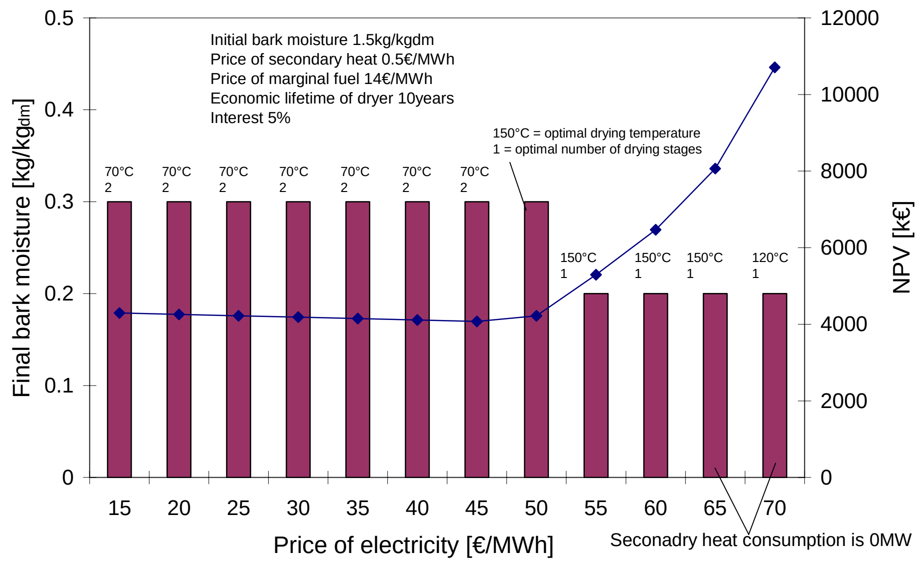


Figure 16. Results of the sensibility study as a function of the electricity price. Reading example: If the electricity price is 40 €/MWh (see other initial values from the figure), the figure shows that the optimal final bark moisture is 0.3 (columns), the optimal number of drying stages 2, the optimal drying temperature 70^{o}C and the net present value approximately 4000k€ (line).

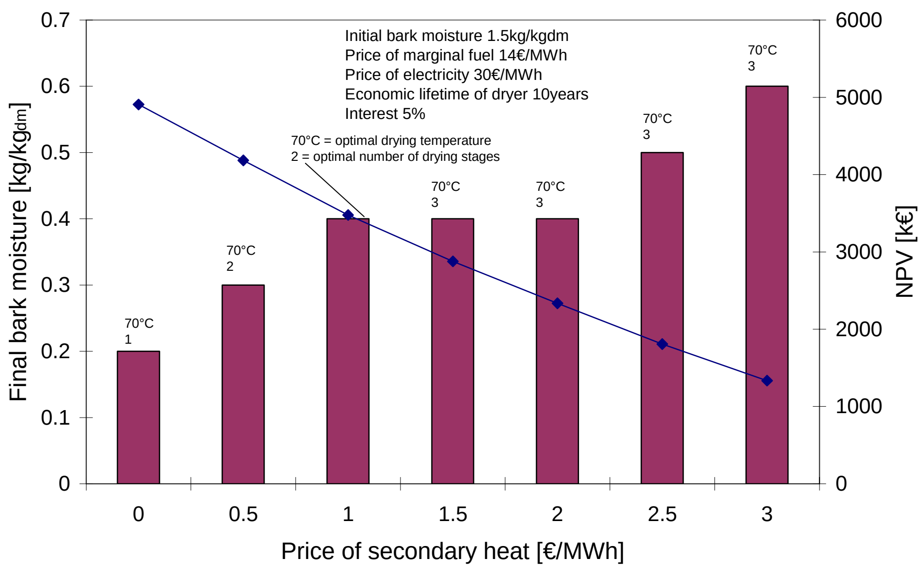


Figure 17. Results of the sensibility study as a function of the secondary heat price. Reading example: If the secondary heat price is 1.5 €/MWh (see other initial values from the figure), the figure shows that the optimal final bark moisture is 0.4 (columns), the optimal number of drying stages 3, the optimal drying temperature 70^{o}C and the net present value approximately 2700k€ (line).

Using the initial values shown in Tables 2 and 3, the optimally designed dryer uses only secondary heat for the heating of drying air, consists of two drying stages and dries the fuel/bark flow to the final moisture content, which is in the order of 0.3kg/kg_{dm} (see Figure 13). The reasons why the results in Figures 14-17 behave as they do are explained in Appendix paper III.

From the viewpoint of the dryer investment the most interesting question is how fuel and electricity prices will develop in the future.

Table 4 shows how the average market prices of the most common mill fuels have developed in Finland between 2001-2005. Table 4 proves the fact that fuel prices have increased considerably in 2005 due to the introduction of the emissions trading (see prices including price of CO_2-ton). Table 4 also reveals that the price of natural gas has increased almost 20 percent from June 2004 to September 2005, even though the effect of emissions trading is excluded. Natural gas is the most important fossil mill fuel used by the forest industry in Finland [5]. Even though it is difficult to predict the price behavior of CO_2-ton merely on the basis of year 2005, it is probable that fuel prices will not decrease considerably in the near future. On the contrary, prices may still increase. If the expected behavior of fuel prices continues, the profitability of the dryer investment improves and the optimal drying system is similar to the one in our basic case (see Figure 13).

At some mills, biofuel may be the marginal fuel part of the year or the whole year, and one may ask whether drying is profitable in these cases. Because of the emissions trading several municipal energy companies have started to replace fossil fuels with biofuels, for example, in district heating, and the market price of biofuel has increased. From the viewpoint of the forest industry, this means that excess biofuel can be sold outside the mill, and drying may still be profitable even if biofuel is the marginal fuel for most of the year. To analyze these cases correctly, the market price of biofuel must be used as a fuel price.

Figure 16 shows that the electricity price has no influence on the profitability and design parameters of the optimal drying system until the price exceeds 45 euro/MWh. The annual average price of electricity has varied between 23 and 36 euro/MWh in the Nordic electricity markets during the years 2001-2005 [49]. In this kind of calculation, it is more reasonable to use the average electricity price over a longer period rather than instantaneous prices. The Nordic electricity markets are characterized by the following factors: (i) Circa fifty per cent of electricity is produced by hydropower and the price usually decreases and increases during rainy and non-rainy years, respectively. (ii) Long cold periods in the winter usually increase the consumption and the price. (iii) Instantaneous electricity prices may be very high due to the reasons mentioned in points (i) and (ii). (iv) Over thirty percent of electricity is produced by nuclear power and renewable energy sources, which are CO_2-free energy production forms. (v) A new nuclear reactor will be introduced in Finland in 2010 and its effect on the future electricity price is unclear. For these reasons it is difficult to estimate the behavior of the average electricity price in the future. On the basis of the average electricity prices during the years 2001-2005, the electricity price does not have any significant influence on the design parameters of the optimal drying system and the profitability of drying. If the average electricity price reaches permanently a clearly higher level in the future, it may have a more important role in the design of the optimal drying system. Finally, it is necessary to state that power-to-heat ratios used in the calculation are relatively high for industrial CHP plants. If the power-to-heat ratios were lower, the electricity price in Figure 16 should be even higher before the optimal design values and the profitability would change.

The values used for economic lifetime and the price of secondary heat are decided by the mill and can vary depending on the actor. On the other hand, these parameters seem merely to influence the final bark moisture content and the number of drying stages.

Table 5. Development of fossil fuel prices in Finland between 2001 and 2005 [50]

	Fuel prices				**Price of CO2-ton**	**Fuel prices including price of CO_2-ton**			
	Coal [€/MWh]	Natural gas [€/MWh]	Peat [€/MWh]	Heavy fuel oil [€/MWh]	[€/t]	Coal [€/MWh]	Natural gas [€/MWh]	Peat [€/MWh]	Heavy fuel oil [€/MWh]
6/2001	12.9	15.8	9.2	25.0					
12/2001	12.4	15.7	9.2	21.8					
6/2002	11.6	15.2	9.2	24.3					
12/2002	11.3	15.7	9.2	25.5					
6/2003	11.0	16.1	9.3	24					
12/2003	11.6	15.6	9.4	23.3					
6/2004	13.4	15.3	9.4	26.2					
12/2004	13.4	16.1	9.7	24.5					
3/2005	13.8	16.2	9.8	29.5	10	17.2	18.2	13.6	32.3
6/2005	13.3	17.0	9.8	33.5	23	21.1	21.6	18.6	39.9
9/2005	13.4	18.2	8.2	39.8	23	21.2	22.8	16.9	46.2

Emission factors for fuels based on lower heating values from IPCC [kg CO_2/MWh fuel used] :
Coal 340.28, Natural gas 201.80, Peat 381.26, Heavy fuel oil 278.26

4.3 Operation of the dryer in an industrial CHP plant

4.3.1 Model description

This chapter presents simulation results of the model that have been developed for analyzing the operation of the dryer integrated into an industrial CHP plant at a pulp and paper mill. The main objective of the analysis is to evaluate whether it is reasonable to equip the dryer with a control system or not.

There is normally considerable variation in the fuel moisture content. The variation is mainly dependent on the season and instantaneous weather. Season and weather also affect the temperature and humidity of the drying air before the first heat exchanger. The created simulation model calculates the final/outgoing fuel moisture content on the basis of the initial fuel moisture content, ambient temperature and humidity.

The availability of process computers has advanced the model-based control with the use of either explicit or knowledge-based models, and the same model built for process simulation can be used for process control [51]. In this connection, the simulation model is also applied to the model-based control of the dryer. *The purpose of the model is to determine drying temperature, or drying temperatures in the case of multi-stage drying, in such a way that the desired final fuel moisture content is always achieved at minimum operating cost (i.e. by maximizing net income/earnings from drying).* The model has been created for a drying system similar to one shown in Figure 7a.

Generally, drying process variables can be divided into input and output/controlled variables [52]. Input variables can be further classified into load and manipulated variables [52]. Load variables (e.g. ambient temperature, humidity) cannot be adjusted by a control system. Manipulated variables (e.g. drying temperature, dry mass flow of fuel) can be adjusted either manually or automatically. In the model, all variables that are possible to manipulate are not adjusted, but have a fixed value, which is the same as the design value. Table 6 shows the group to which each variable belongs in this model. If the variable that is possible to manipulate is not adjusted, it is called a fixed parameter in Table 6.

Table 6. Fixed, load, controlled and manipulated variables in the simulation model

Parameter/variable	Symbol	Group of variable	Explanation
Air mass flow	$\dot{m}_{da}$	Fixed	Same as the design value
Particle size	-	Fixed	Same as the design value
Material properties of bark	-	Fixed	Same as the design values
Cross-sectional area of the dryer	A	Fixed	Same as the design value
Bed height	Z	Fixed	Same as the design value
Dry mass flow of bark	$\dot{M}_{dm}$	Fixed	Same as the design value
Initial bark moisture content	u_{in}	Load	Depends on material, weather and season
Outdoor humidity	x_1	Load	Depends on weather and season
Outdoor temperature	t_1	Load	Depends on weather and season
Final bark moisture content	u_{out}	Controlled	Same as the desired set value
Drying temperatures	t_4, t_8, t_{12}…..	Manipulated	Model calculates optimal drying temperature(s)
Heat inputs	Φ_{sh} ,Φ_{bs} ,Φ_{es}	Manipulated	Drying temperatures are adjusted to optimal ones by manipulating heat inputs

Table 6 shows that fuel and air input (dry mass flows of fuel and air) into the dryer are assumed to be fixed and manipulated parameters are drying temperatures, which are determined in each drying stage by the model. To be precisely, the actual parameters to be manipulated are heat inputs in to the dryer (also shown as a manipulated parameter in Table 6), which are adjusted by conventional controllers in order to achieve calculated drying temperatures. Because the model calculates drying temperatures, they are called manipulated parameters instead of heat inputs.

The simulation model is presented in detail in Appendix paper IV and is not discussed any further here. To apply the model to model-based control the initial fuel moisture content must be known on-line. A conceivable method to approximately determine the initial fuel moisture on-line is discussed in the next chapter and the simulation results are presented in chapter 4.3.2.

4.3.2 Method for on-line determination of initial fuel moisture content

The most reliable way to determine the initial fuel moisture content is the drying of fuel samples in an oven. At least two samples are taken from the fuel flow and kept in an oven at 105±2°C for at least 16 hours but not longer than 24 hours [53]. After the drying, it can be assumed the samples are completely dry and the initial moisture content calculated on wet basis is defined as follows:

$$w_o = \frac{m_1 - m_2}{m_1}\,, \tag{18}$$

where m_1 is the mass of the sample before drying and m_2 after drying. Although the method is reliable and simple, it is time consuming and it must be carried out manually, which is not desirable.

Appendix paper V presents a simple method for approximately determining the initial fuel moisture content. The method is based on the on-line measurement of the outlet air moisture at the beginning of drying. The idea of the method is to find a correlation between fuel moisture content and the change in the outlet air moisture content. The idea was experimentally tested by drying regular shaped wood particles (20x20x5mm) in a test rig (see Figure 3). The evaporation surface in each test was the same but the moisture contents of the particles were different. Figure 18 shows the change in the outlet air moisture for different initial moisture contents of particles and also for particles which were frozen before drying.

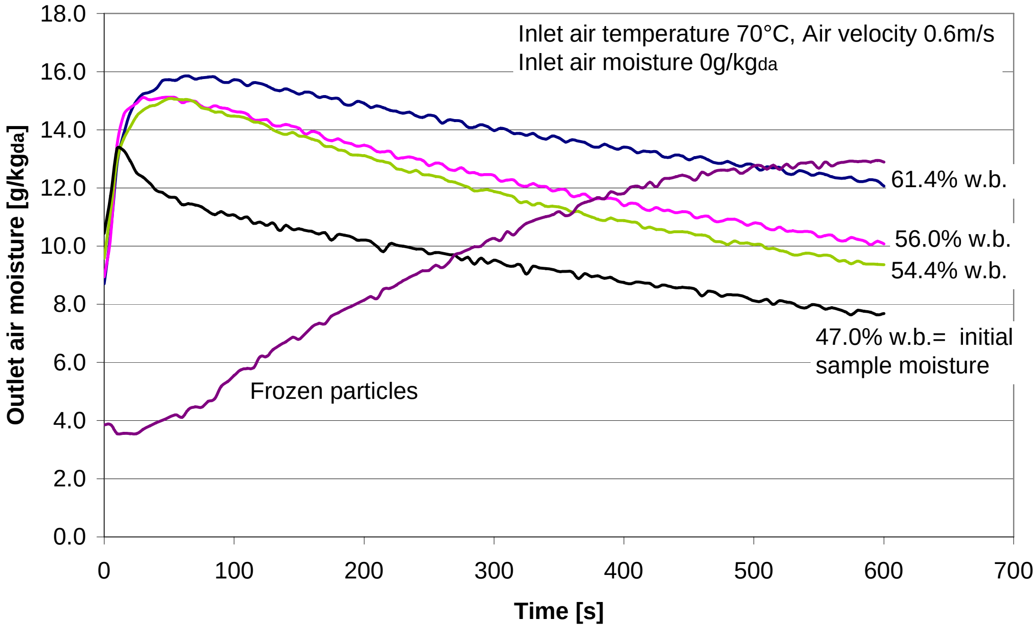


Figure 18. Change in outlet air moisture content for different initial moisture contents of sample particles and for frozen particles.

Figure 18 shows that there is a relatively clear correlation between the change in outlet air moisture content and the initial moisture content of the particles. Even a relatively small moisture difference between particles (56% and 54.4 %) can be observed. Appendix paper IV proposes that outlet air moisture content is measured in the drying chamber after the fuel has entered the chamber. This is one possible way to carry out the measurement but probably not the best way. A more comprehensive way is to take a small sample from the fuel flow to a separate test chamber before the drying chamber. By measuring the outlet air moisture content in a separate chamber with constant volume the evaporation surface is approximately the same in each measurement, which improves the accuracy of the determination. In addition, the bed height in a separate chamber can be clearly lower than in the drying chamber, in which case the change of outlet air moisture content is observed faster.

To ensure the suitability of the method in the case of a commercial dryer, measurements for actual biofuels with varying particle size must be done. Instead of outlet air moisture content, measurement of outlet air temperature may give sufficient information for the determination of initial fuel moisture content, and should be experimentally tested, too.

4.3.3 Results and discussion

The model has been applied to a biofuel drying study where a virtual dryer is integrated into an actual CHP plant. The process values of the CHP plant are the same as in the case study in Chapter 4.2 (see Table 2). Chapter 4.2 concludes that the optimally designed dryer uses only secondary heat as heating energy when it operates at its design values. Table 7 shows design values for four dryers which are designed to use secondary heat as a heat source. In simulations, these dryers are first equipped with a model-based control and then the results are compared with similar dryers with no control.

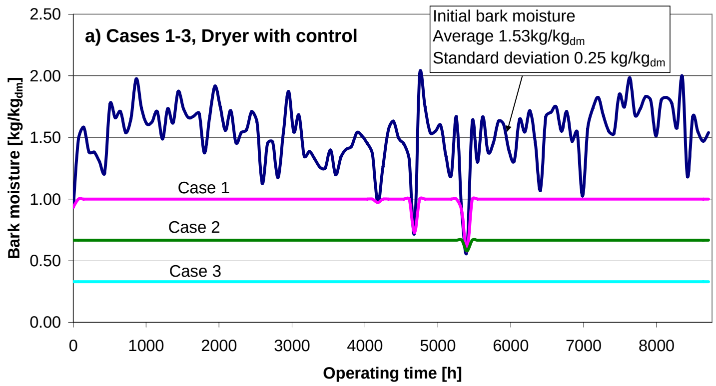
The material to be dried is soft wood (pine) bark. The load variables (bark moisture contents, outdoor temperatures, and humidity values) for the drying simulations are obtained from the same mill where the CHP-plant is situated (see load variables in Appendix paper V). In the simulations, the load variables have 122 values and they cover the whole calendar year (load values change every 72 hours). In practice, the inlet values may change more frequently. The electricity prices used in the calculations change every month. The prices are based on data from the Nordpool for the period the beginning of 2000 to the end of 2003 [49]. The prices of marginal fuel and secondary heat are 10€/MWh and 0.5 €/MWh, respectively.

Table 7. Design values of dryers in simulation cases

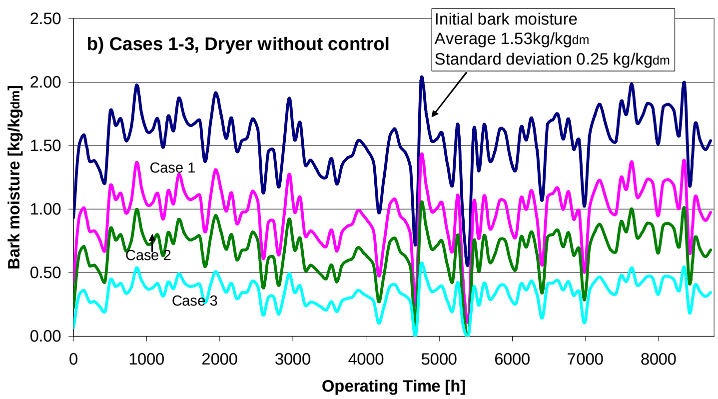
Case	Initial bark moisture content [kg/kg$_{dm}$]	Final bark moisture content [kg/kg$_{dm}$]	Number of drying stages	Drying temperature [^{o}C]	Residence time [s]
1	1.53	1.00	1	70	1035
2	1.53	0.667	1	70	1695
3	1.53	0.333	1	70	2730
7	1.53	0.667	3	70 (in each stage)	1895
Design values similar in all cases: Initial bark moisture content 1.53 kg/kg$_{dm}$, bed thickness 200mm, air velocity per free sectional area 0.65m/s, outdoor temperature 5^{o}C, humidity 0.004kg/kg$_{da}$, , effectiveness for secondary heat exchanger 0.89, effectiveness for backpressure steam exchanger 0.73, effectiveness for extraction steam exchanger 0.67					

The results for the simulation cases are shown in Figures 19-20. The figures show final bark moisture contents and annual net incomes of drying for the dryers equipped with the model-based control and for the dryers with no control. Net incomes of drying are expressed as euro/t$_{dm}$. The simulation results are independent of the dry mass flow and the net incomes are directly proportional to dry mass dried in the dryer.

Appendix paper V also discusses cases where the dryer is designed to use both secondary heat and backpressure steam as heat sources. On the basis of our results in Chapter 4.2.2, these cases are not relevant.

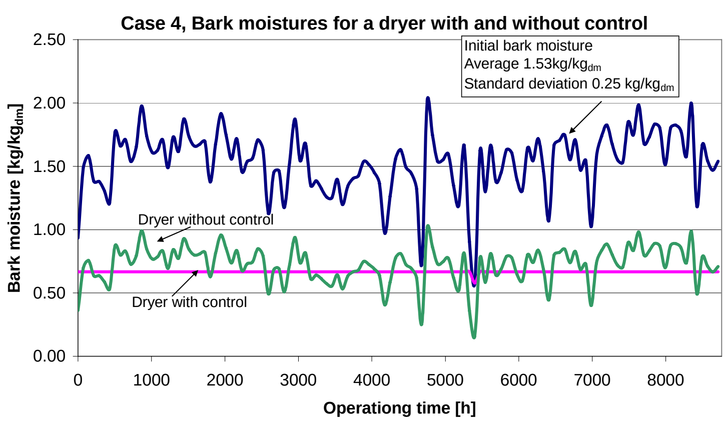


	Case1	Case2	Case3
Average final moisture content, kg/kg$_{dm}$	0.992	0.666	0.330
Standard deviation, kg/kg$_{dm}$	0.046	0.008	0.000
Net income, €/t$_{dm}$	2.746	4.442	6.004



	Case1	Case2	Case3
Average final moisture content, kg/kg$_{dm}$	0.971	0.675	0.344
Standard deviation, kg/kg$_{dm}$	0.222	0.181	0.107
Net income, €/t$_{dm}$	3.325	5.024	6.792

Figure 19. Behavior of final bark moisture content and net income of drying in cases 1-3 (see cases from Table 7) on the basis of simulation calculations: a) dryer with control b) dryer without control.



	Dryer with control	Dryer without control
Average final moisture content, kg/kg$_{dm}$	0.666	0.642
Standard deviation, kg/kg$_{dm}$	0.008	0.129
Net income, €/t$_{dm}$	4.784	5.348

Figure 20. Behavior of final bark moisture content and net income of drying in case 4 (see case from Table 7) on the basis of simulation calculations: a) dryer with control b) dryer without control (Note case 4 is case 7 in Appendix paper V).

The simulation results show that it is possible to stabilize the final bark moisture content by controlling the drying temperatures. The results also show that the deviation in final bark moisture content decreases even when the dryer has no control. The lower the designed value for the final bark moisture content, the more the deviation decreases.

The net income from the drying increases, when the dryer has no control. There are two reasons for bigger the net income when the dryer has no control. Firstly, the whole potential of affordable secondary heat is always used in drying. In the case of the control, the whole potential is not used if the initial fuel moisture content is below the design value. Secondly, the dryer never uses more expensive steam in drying. In the case of control steam is consumed, if the initial fuel moisture content is above the design value. On the basis of the assumptions made in the case study, the need for a control cannot be justified on economic grounds. However, there may be a need for a control, if the use of homogenous, and not only dry, fuel has any notable influence on the following issues:

- Operation and control of the boiler
- Boiler efficiency
- Predictive control of the power plant in cases where the heat consumption at the mill changes.
- Deviations in (live or process) steam pressure resulting from changes in heat consumption at the mill.

An analysis of the above mentioned issues is beyond the scope of this study but should be investigated to be able to answer ultimately the question about the need to control the final fuel moisture content. If the use of homogenous fuel has positive influences on some of the above mentioned issues, the control may have economic grounds, too.

The answer is also slightly dependent on whether the dryer is integrated into an existing or a new boiler. Existing fluidised bed boilers have been designed for moist and heterogeneous fuels. In particular, bubbling fluidised bed boilers (BFB boiler) may not even tolerate too dry fuel, which can cause, for example, ash melting as a result of the excessively high combustion temperature. To avoid the use of too dry fuel in the boiler, the dryer may need control. On the other hand, mixing of moist fuel with dry fuel after the dryer may be an alternative method to avoid the use of too dry fuel in the boiler. Otherwise the need for control depends on analysis results of the previously mentioned issues. In the case of a new boiler, the boiler can be already designed for dry and homogenous fuel (i.e. fuel with a constant heating value) if the dryer is equipped with a control for some reason. The model presented in Appendix paper V is a conceivable method of carrying out the model-based control of the dryer.

4.4 Conclusions on integration and operation of the dryer

The following main conclusions can be drawn concerning integration and operation of the dryer at industrial CHP plant:

- Using current energy prices and an economic lifetime of 10 years with an interest rate of 5%, the optimally designed dryer uses only secondary heat for the heating of drying air, consists of two drying stages and dries the fuel/bark flow to a final moisture content which is in the order of 0.3kg/kg_{dm} (see Figure 13).
- If fuel prices are expected to rise further in the future, the profitability of the dryer investment improves and the optimal drying system is similar to the one in our basic case (see the first bullet).
- The electricity price has not any significant influence on the design parameters of the optimal drying system and the profitability of drying in the case of study until the annual average price exceeds 45-50 euro/MWh in the case study (see Figure 16).
- Predicting future electricity prices is difficult in the Nordic electricity markets, and the rule of thumb is that expected earnings from drying primarily stem from decreased marginal fuel consumption, and not increased power generation.
- On the basis of the assumptions made in the simulation case study, there are no economic grounds for homogenizing the final fuel moisture content by means of control.
- However, the use of homogenous fuel may have some positive influences on the boiler and the power plant process, which may favour the introduction of control. To be able to ultimately answer the question about the need to control the final fuel moisture content, these influences need to be analysed in collaboration with power plant specialists.
- If the dryer is equipped with control it can be carried out as a model-based control using the model presented in Appendix paper V.

5 Evaluation of energy efficiency in drying

The main goal of this chapter is to present two methods for evaluating the energy efficiency in drying. The evaluation methods are the specific heat consumption and the irreversibility rate. Both evaluation methods are applied to two flow sheets, which are alternative methods of decreasing the heat consumption in drying. The flow sheets considered in this chapter are multi-stage drying (MSD) and single-stage drying with partial recycling of drying air (SSD). MSD is the primary method of improving the energy efficiency of drying in this thesis. SSD in introduced as an alternative flow sheet for improving the energy efficiency. To further decrease the heat consumption, both drying systems are provided with an exhaust air heat recovery unit.

5.1 Single- stage drying with partial recycling of drying air

Figure 21a shows a continuous single-stage dryer with partial recycling of spent air including a heat recovery unit. Figure 21b shows also a multi-stage drying system with a heat recovery unit. Multi-stage drying is discussed in more detail Chapter 3.1. The changes of the states of the drying air on an enthalpy humidity chart in single-stage drying with partial recycling of air is illustrated in Figure 22. Partial recycling of the spent drying air and the heat recovery reduce heat consumption in drying process, because the air temperature before the heating increases. Because of the recycling, the fresh drying air has a lower potential for taking up moisture, which leads to a bigger air mass flow through the drying chamber.

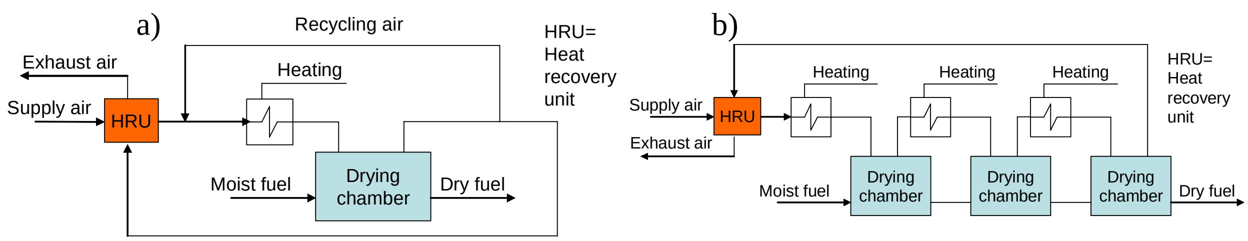
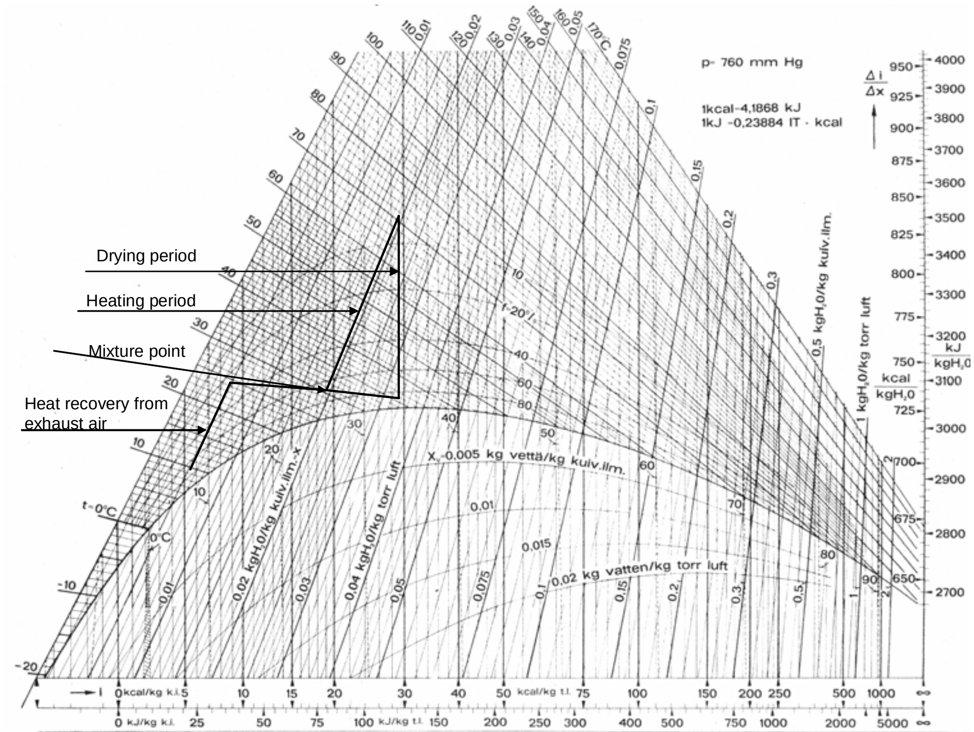


Figure 21. Methods of improving energy efficiency a) Partial recycling of spent drying air+ heat recovery (SSD) b)Multi-stage drying + heat recovery (MSD).



Translations:
kuiv. ilm.= *dry air*
torr luft = *dry air*
kg vettä/ kg kuiv.ilm. = *kg water / kg dry air*
kg vatten / kg toff luft = *kg water / kg dry air*

Figure 22. Changes of the states of the drying air on an enthalpy humidity chart in single-stage drying with partial recycling of air.

5.2 Evaluation methods

Traditionally, the most common way to evaluate the energy efficiency in drying is the specific heat consumption [54]:

$$q = \frac{\sum_{i=1}^{n} \Phi_i}{\dot{m}_e} \quad , \qquad (19)$$

where n is the number of drying stages, ϕ the heat input into the drying system, and $\dot{m}_e$ the evaporation rate of the dryer.

Equation (19) is based on the energy balance and it does not pay any attention to the temperature level at which drying occurs. If the dryer is integrated into a power generation process the drying temperature is not irrelevant from the perspective of energy efficiency. To take into account the effect of temperature, the evaluation method must be based on the Second Law of Thermodynamics. The term irreversibility rate is presented in [55] and is defined as follows:

$$I = T_o [\sum_{i=1}^{n} m_i \Delta s_i - \sum_{i=1}^{m} \frac{\Phi_i}{T_{ri}}] \quad , \tag{20}$$

where T_o is the temperature of the surroundings, Δs the change of entropy for each stream of matter through the system, and Φ_i the heat inputs and outputs to the system. In this paper, the heat input to the system is taken as positive. T_r is the temperature at the point on the boundary of the system where the heat transfer is taking place [55]. Generally, the term Φ_i/T_i is known as a thermal entropy flux [55].

If the steam in a turbine can expand to the pressure corresponding to the temperature of the surroundings, the exergy balance for a steam power plant with a dryer inside the control region is [55]:

$$E_f + E_a - E_P = W + I_d + I_x \quad , \tag{21}$$

where E_f is the exergy of the fuel, E_a the exergy of the air entering the control region, E_p the exergy of the streams leaving the control region, W the mechanical work, I_d the irreversibility rate of the drying and I_x the irreversibility rates of the other sub-processes. It can be assumed that E_p and I_x remain almost the same regardless of the drying process. The difference between the irreversibility rates of two drying processes gives how much mechanical work (electricity) will be lost if the dryer with the higher irreversibility rate is used instead of the dryer with the lower rate. However, it is necessary to state that maximizing the power generation is usually not the main aim in a CHP-production but the main aim is to produce heat for manufacturing processes.

Determination of the heat consumption and irreversibility rate for MSD and SSD are explained in detail in Appendix paper VI.

5.3 Results and discussion

Equations (19) and (20) have been applied to a drying case for evaluating the energy efficiency of the drying process. In the drying case, moist biofuel is dried from an initial moisture content of 1.5kg/kg_{dm} to a final moisture content of 0.3kg/kg_{dm}, and the dry mass flow of the fuel is 1kg_{dm}/s. Two alternative heat sources are available for the heating: steam at a pressure of 3bars (133^{o}C), and water at a temperature of 80^{o}C. The minimum temperature difference between air leaving the heat exchanger and the heat source entering the heat exchanger is 5^{o}C, in which case the inlet air temperatures of the drying chambers are 128^{o}C and 75^{o}C. The outlet temperature of water is assumed to be 10^{o}C higher than the inlet air temperature of the heat exchanger. In the heat recovery unit, the minimum temperature difference is 5^{o}C. The outdoor temperature is 15^{o}C and the absolute moisture content 0.0064kg/kg_{da} corresponding to a relative humidity of 60%. Mechanical work inputs into the dryer and the heat losses are neglected. Neither are there any indirectly supplied heat inputs into the drying chamber. The drying air is assumed to be fully saturated after the drying chamber.

Figure 23 shows heat consumptions and the irreversibility rates for multi-stage drying as a function of drying stages. Heat consumptions and the irreversibility rates for SSD are presented in Appendix paper VI.

To compare the heat consumption and irreversibility rate between MSD and SSD, the dryer dimensions are altered so that they are the same. It is assumed that the dryer dimensions are proportional to the cross-sectional area of the dryer, which depends on the air mass flow as follows:

$$A = \frac{\dot{m}_a}{\rho_{da} v_a} \quad , \tag{22}$$

The term $\dot{m}_a$ is defined as the product of the number of drying stages and the mass flow of dry air through the multi-stage dryer ($\dot{m}_a = n\dot{m}_{da}$). The mass flow $\dot{m}_a$ is first determined for the MSD, because the number of drying stages must be an integer. In SSD, the recycle ratio is given such a value that the air mass flow through the drying chamber is the same as $n\dot{m}_{da}$. Instead of the air mass flow or the cross-sectional area, the energy consumption and the irreversibility rate are expressed as a function of a dimensionless value, defined as follows:

$$\frac{A}{A_{ref}} = n \frac{\dot{m}_{da}}{\dot{m}_{aref}} \, , \tag{23}$$

where A/A_{ref} is the ratio of dryer areas and $\dot{m}_{aref}$ the mass flow of the dry air in single-stage-drying with no recycling (R=0 and n=1). This is the smallest dryer, and the value of RDA is always bigger than 1 when the number of drying stages increases. In the example calculations, the maximum number of drying stages is reached when the recycle ratio is approximately 0.9. Figure 24 shows the comparison results for the inlet air temperature 75°C, in which case air is heated using secondary heat. Comparison results for the inlet air temperature 128°C are presented in Appendix paper VI.

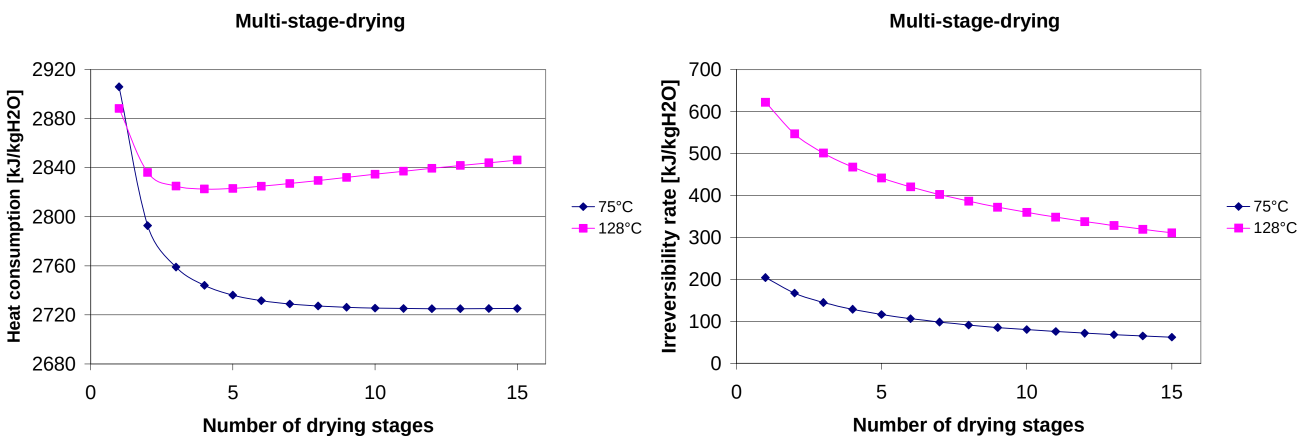


Figure 23. Heat consumptions and irreversibility rates for multi stage drying system including heat recovery from exhaust air (T_o=15^{o}C).

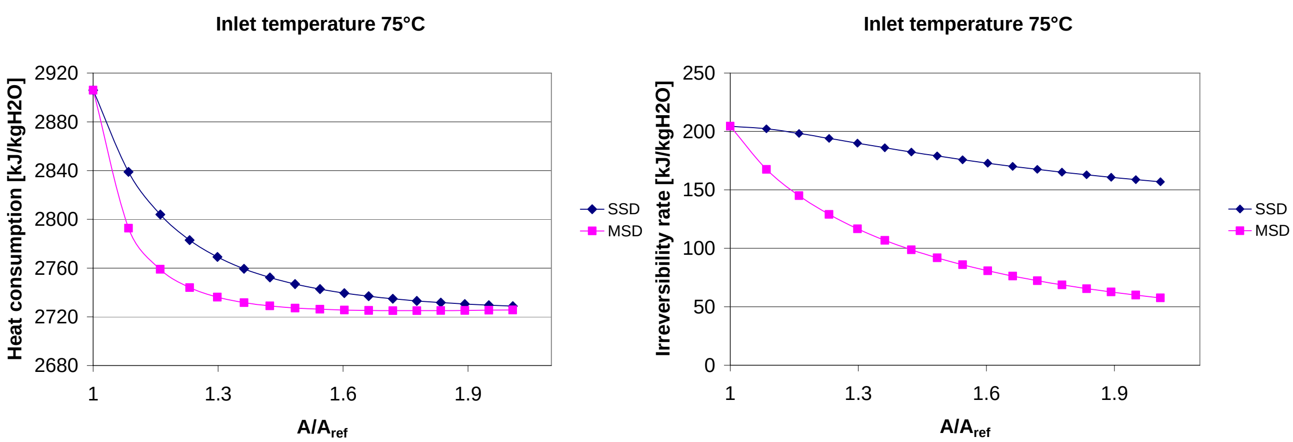


Figure 24. Comparison of energy consumption and irreversibility rate between single-stage drying with recycling (SSD) and multi-stage drying (MSD) for an inlet temperature of 75^{o}C (T_o=15^{o}C).

Figure 23 shows that the drying temperature has little effect on heat consumption. The difference is in the order of 100kJ/kg$_{H2O}$. In stead, the irreversibility rate clearly depends on the drying temperature. The differences between irreversibility rates are several hundred kilojoules per mass of water evaporated. When steam is used, heat transfer into the drying system occurs at a higher temperature level than in the case of secondary heat. The average temperature difference between the heat source and air is also higher for steam, because the steam temperature remains constant. Heat transfer occurring over a high temperature difference also increases the irreversibility rate. The difference between the irreversibility rates gives theoretical losses in electricity production, which is more relevant information than the difference between heat consumptions when the dryer is integrated into a power generation process. However, it is important to consider that the irreversibility rate gives only a theoretical estimate of losses. In practice, it is difficult to utilise secondary heat in electricity production.

The calculated values in Figure 23 are very low because the heat losses and the pressure drops of the drying system are neglected, the drying air is assumed to be fully saturated after the drying chamber, and the drying system is provided with a heat recovery unit. Only the increase in fuel temperature and temperature differences in the heat exchangers have been considered. The calculated heat consumptions and irreversibility rates are good reference values for the estimation of the energy efficiency of real air drying systems.

The comparison between the drying systems with the same dimensions (Figure 24) shows that the heat consumption is almost the same for both systems. The difference varies from 0 to 50kJ/kg$_{H2O}$. However, the irreversibility rate is smaller for MSD than for SSD. The main reason for the lower irreversibility rate is the entropy change, which becomes smaller for MSD. According to these results, multi-stage drying seems to be a slightly more energy efficient drying system than single-stage drying with partial recycling of the drying air.

5.4 Conclusions on evaluation of energy efficiency

- By means of various flow sheets and heat recovery from exhaust gas the lowest possible heat consumptions in air drying are close to 2700 kJ/kg$_{H2O}$.
- If the energy used in drying can be converted into mechanical work, the evaluation of energy efficiency should be based on exergy rather than on energy analyse.
- The irreversibility rate is a conceivable method for evaluating the energy efficiency of drying processes from the viewpoint of the Second Law.
- To decrease the value of the irreversibility rate, the drying temperature should be as low as possible, and the heat transfer should occur over small temperature differences.
- According to the results shown in Chapter 5.3, multi-stage drying seems to be a slightly more energy-efficient drying system than single-stage drying with partial recycling of spent drying air, when the dryer dimensions are the same.

6 Concluding remarks

The integration and operation of the dryer at an industrial CHP plant has been analyzed in a case study. According to the case study results, an optimally designed dryer uses only secondary heat, not steam, as a heat source, and earnings stem from decreased marginal fuel consumption, not increased power generation. If fuel prices are assumed to increase in the future, as seems likely, the optimal design parameters of the dryer are similar to those in a basic case and the profitability of the dryer investment improves. However, it is necessary to state that the profitability of the dryer investment depends always on the actor and the case and must be assessed separately in each case. Guidelines for the assessment of the profitability of the dryer investment are presented in this thesis. At the moment, the prospects for profitable dryer investments are good.

On the basis of the assumptions made in the simulation calculations, there does not seem to no economic grounds to control the final fuel moisture content by adjusting heat inputs into the dryer. However, the use of homogenous, and not only dry, fuel may have positive influences on the operation/control of the boiler, boiler efficiency and predictive control of the power plant when heat consumption changes at the mill. To be able to answer ultimately the question about the need to control the final fuel moisture content, these questions should be analysed. If the dryer is equipped with control it can be carried out as a model-based control using the model presented in Appendix paper V.

Experimental tests show that critical moisture content is high for woody-based fuels and it is necessary to use diffusion-based drying models to determine drying the times theoretically in a fixed bed. Drying times calculated by the constant drying model are too inaccurate compared to actual drying times, even though the final fuel moisture content was clearly higher than the fiber saturation point. From viewpoint of both constant and diffusion-based drying models, the extremely wide particle size distribution in biofuel flow also sets challenges for the development of theoretical drying models. In this study, drying times used in the calculation models have been determined experimentally in a fixed bed batch dryer. Experimentally determined drying times are analogous to a continuous cross-flow process.

Guidelines for dimensioning the conveyor/fixed bed dryer in the case of multi-stage drying have bee presented. In the design of the dryer, air velocity must be slightly lower than the minimum fluidized velocity and the bed thickness must be at least so high that the drying air reaches its saturation point at the beginning of the drying. Determining the dryer dimensions in the case of multi-stage drying is based on the use of the wet bulb temperature and fuel flow sharing.

Specific heat consumption is the most common method of evaluating the energy efficiency in drying. The weakness of specific heat consumption as an evaluation method is that it does not pay any attention to the drying temperature. On the basis of the calculated results the drying temperature has a quite small effect on the calculated heat consumptions. To take into account the effect of the drying temperature on the energy efficiency an alternative evaluation method, i.e. the irreversibility rate, has been introduced. The results show that there is a clear difference between irreversibility rates depending on the drying temperature. If the energy used in drying can be converted into mechanical work, irreversibility rate is a more comprehensive method of comparing energy efficiency between different drying processes than specific heat consumption.

7 Recommendations for the future research and development of the dryer

Building a pilot-dryer is a necessary step before commercialization of the dryer. Robustness is one of the key requirements for a commercial dryer and it is impossible to evaluate the robustness reliably without experiences from a pilot-dryer. Fuel input and output, dust formation, blockage risk, fire risk, fouling, corrosion, the need for maintenance as well as moisture distributions in vertical and horizontal directions are the main issues that must be analysed by means of the pilot-dryer. A pilot-dryer is also the most reliable way to specify presented design parameters.

The use of dry and homogenous fuel may have positive influences on the operation and control of the boiler and power plant process. The main issues to be analysed in this connection are listed in Chapter 4.3.3 but have not been investigated any further in this study. Especially, from the viewpoint of controlling the final fuel moisture in the dryer it would be important to analyse these issues. If the control of the final fuel moisture content entails clear benefits compared with the uncontrolled dryer, the idea of the model-based control must be implemented in the pilot-dryer. Both the boiler analysis and the possible implementation of the model-based control must be carried out in collaboration with boiler and control specialists.

References

[1] The Finnish Bio Energy Society. Bioenergy terms. http://www.finbioenergy.fi/default.asp?init=true&InitID=398;0, February 28, 2006.

[2] The Finnish Ministry of Trade and Industry. Renewable energy sources and peat. http://www.ktm.fi/index.phtml?l=en&s=179, February 28, 2006.

[3] Alakangas E. Properties and standadization of solid fuels. In: Alakangas E, Agar D, Leppänen M, editors. Report of the seminar presentations and study tour. Jyväskylä, Finland: OPET CHP/DHC; 2003.

[4] Vanhatalo J, Rautanen J, Ahtila P. Metsäteollisuuden energiantuotannon hiilidioksidipäästöjen vähentäminen biomassoilla (Reduction of the carbon dioxide emissions by biomasses in the energy production of the forest industry). Publications of Laboaratory of Energy Economics and Power Plant Engineering, Raport A2. Espoo, Finland: Helsinki University of Technology; 1999.

[5] Finnish Forest Industries Federation. Finnish Forest Industries Yearbook 2005, Statistics 2004. Helsinki, Finland: Finnish Forest Industries Federation; 2005.

[6] Statistics Finland. Energiatilastot 2004 (Energy Statistics 2004). http://www.tilastokeskus.fi/til/ekul/index.html, February 28, 2006.

[7] Huhtinen M, Hotta A. Combustion of Bark. In: Gullichsen J, Fogelholm C-J, editors. Chemical Pulping. Helsinki, Finland: Papermaking Science and Technology; 1999. p. 203-301.

[8] Alakangas E. Suomessa käytettävien polttoaineiden ominaisuuksia (Properties of fuels used in Finland). Research Notes of Technical Research Centre of Finland. Espoo: Technical Research Centre of Finland; 2000.

[9] Schulze WJ. Combustion Control and Burner Management. In: Elliot TC, Chen K, Swanekamp RC, editors. Standard Handbook of Powerplant Engineering. New York: McGraw Hill; 1997. p. 5.33-5.45

[10] Jegoroff M, Polttoaineen syöttö ja sen vaikutukset leijukerrospoltossa (Fuel input and its effect on fluidized bed combustion). In: Koskelainen L, Roiha I, editors. Proceedings of Kattilapäivät seminar. Finland: Lappeenranta University of Technology; 2004.

[11] Joronen T. Bio- ja seospolttoaineiden syötön ja lämpöarvon hallinta automation avulla. (Control of fuel input and heating value of bio- and mixture fuels by means of automation). In: Koskelainen L, Roiha I, editors. Proceedings of Kattilapäivät seminar. Finland: Lappeenranta University of Technology; 2004.

[12] Wimmerstedt R. Recent advances in biofuel drying. Chemical Engineering and Processing 1999; 38: 441-447.

[13] Pellikka J, Saviharju K, Kuoren ja puujätteen poltto (Combustion of bark and waste wood). In: Virkola NE, editor. Puumassan valmistus osa 2 (Pulp making part 2). Finland: Suomen paperi-insinöörien yhdistys; 1983. p. 1519-1582.

[14] Wimmerstedt R, Hallsröm A. Torkning av torv och skogsbränslen: teknik, ekonomi, utvecklingsbehov (Drying of peat and biofuels: Techniques, economy, development needs). Lund University Report, LUTKDH/ TKKA-3002. Lund, Sweden: Lund University; 1984.

[15] Tapanainen J. Turpeen ja puubiomassan kuivaus: kirjallisuuskatsaus (Drying of peat and woody-based biomasses: Literature review). Research Notes of Technical Research Centre of Finland. Espoo, Finland: Technical Research Centre of Finland; 1982.

[16] Linna V, Järvinen T. Hakkeen keinokuivauksen tekniset ratkaisut ja taloudellisuus (Technical solutions and economy of the artificial drying of wood chips). SITRA Research Report No. 25 A. Helsinki, Finland: SITRA, The Finnish Ministry of Trade and Industry; 1983.

[17] Magnusson L. Uppföljning av erfarenheter från anläggningar för torkning av biobränsle (Follow-up of operation experiences of biofuel dryers). Värmeforskrapport 479. Stockholm, Sweden: Värmeforsk; 1993.

[18] Johansson I, Larsson S, Wennberg O. Torkning av biobränslen med spillvärme (Drying of biofuels utilizing waste heat). Värmeforskrapport 881. Stockholm, Sweden: Värmeforsk; 2004.

[19] Berge C, Dejfors C. Drifterfarenheter från ångtorkar och direkta rökgastorkar (Operation Experiences of Biofuel Dryers). Värmeforskrapport 681. Stockholm, Sweden: Värmeforsk; 2000.

[20] Berghel J, Renström R. Basic design criteria and corresponding results performance of a pilot-scale fluidized superheated atmospheric condition steam dryer. Biomass&Bioenergy 2003; 23:103-112.

[21] Hovmand S, Fluidized bed drying. In: Mujumdar AS, editor. Handbook of industrial drying. New York: Marcel-Decker;1995. p. 195-248.
[22] Wimmerstedt R. Drying of peat and biofuels. In: Mujumdar AS, editor. Handbook of industrial drying. New York: Marcel-Decker; 1995. p. 809-859.
[23] Fagernäs L, Sipilä K. Emissions from Biomass Drying. Proceedings of Conference on Developments in Thermochemical Biomass Conversion. Banff, Canada; 20-24 May, 1996.
[24] Spets JP, Ahtila P. Reduction of organic emissions by using a multistage drying system for wood-based biomass. Drying Technology: 2004; 22: 541-561.
[25] Parviainen K, Fogelholm C-J, Manninen J. Secondary heat systems and pinch technology. In: Gullichsen J, Fogelholm C-J, editors. Chemical Pulping. Helsinki, Finland: Papermaking Science and Technology;1999. p. 336-374.
[26] Mujumdar AS, Menon AS. Drying of Solids: Principles, Classification, and Selection of Dryer. In: Mujumdar AS, editor. Handbook of Industrial Drying. New York: Marcel-Decker;1995. p. 1-39.
[27] Spets J-P. Enhancement of The Use of Wood Fuels in Heat and Power Production in Integrated Pulp And Paper Mill. Doctor Thesis. Espoo, Finland: Helsinki University of Technology; 2003.
[28] Olsson F, Hansson H, Egard M, Bärring M. Tillvaratagande av spillvärme (Utilization of secondary heat). Värmeforskrapport 824. Stockholm, Sweden: Värmeforsk; 2003.
[29] Ahtila P, Svinhufvud P. Heat Recovery from Secondary Heat Flows of the UPM-Kymmene Kaukas Mill. European Conference on Industrial Energy, Vienna; 1998.
[30] Kilponen L, Ahtila P, Parpala J, Pihko M. Improvement of pulp mill energy efficiency in an integrated pulp and paper mill. Research report TKK-EVO-A5. Espoo, Finland: Helsinki University of Technology; 2000.
[31] Renström R, Berghel J. Drying of Sawdust in an Atmospheric Pressure Spouted Bed Steam Dryer. Drying Technology 2002; 20: 449-464.
[32] Berghel J, Renström R. Controllability of Product Moisture Content When Non-screened Sawdust is Dried in a Spouted Bed. Drying Technology 2004; 22: 507-519.
[33] Axby F. Styrning av biobränsletorkar (Control of biomass dryers), Värmeforskrapport 671. Stockholm, Sweden: Värmeforsk; 1999.
[34] Nyström J, Dahlqvist E. Methods for determination of moisture content in woodchips for power plant – a review. Fuel 2004; 83: 773-779.
[35] Impola R, Saastamoinen J, Fagernäs L. Kiinteiden polttoaineiden kuivumisparametrien karakterisointi (Characterization of drying parameters for solid fuels). Final Report of Bioenergy Research Programme-301. Jyväskylä, Finland: Technical Research Centre of Finland; 1996.
[36] Saastamoinen J, Impola R. Drying of Solid fuel Particles in Hot Gases. In: Rudolph V, Keey RB, Mujumdar AS, editors. Proceedings of the 9th International Drying Symposium, Gold Coast, Australia; 1-4 August, 1994. p. 623-630.
[37] Saastamoinen J, Impola R. Drying of Biomass Particles in Fixed and Moving Beds. In: Strumillo C, Pakowski Z, Mujumdar AS, editors. Proceedings of the 10th International Drying Symposium, Krakow, Poland; 30 July-2 August, 1996. p. 349-356.
[38] Impola R, Saastamoinen J, Kääriäinen J. Characterization of Drying Properties of Biofuels. In: Chartier PH, Beenackers AACM, Grassi G, editors. Proceedings of the 8th European Biomass Conference, Vienna, Austria; 3-5 October, 1994. p. 1498-1505.
[39] Johasson A, Fyhr C, Rasmusson A. High temperature convective drying of wood chips with air and superheated steam. International Journal of. Heat and Mass Transfer 1997; Vol. 40 No.12: p 2843-2858.
[40] Perre P, Morse M, Martin M. Advances in transport phenomena during convective drying superheated steam and moist air. International Journal of Heat and Mass Transfer 1993; Vol. 36: No. 11: p. 2725-2746.
[41] Wang ZH, Chen G. Heat and mass transfer in fixed-bed drying. Chemical Engineering Science 1994; 54: p. 4233-4343.
[42] Keey RB. Drying: Principles and Practice. International series of monographs in chemical engineering, 13. Oxford; 1972.
[43] Lydersen AL. Mass Transfer in Engineering Practice. Chichester, New York, Brisbane, Toronto, Singapore: John Wiley&Sons; 1983.

[44] Lampinen MJ. Kemiallinen termodynamiikka energiatekniikassa (Chemical Thermodynamics in Energy Engineering). Publications of Laboratory of Applied Thermodynamics, Espoo, Finland: Helsinki University of Technology; 1996.

[45] Hyppänen T, Raiko R. Leijupoltto (Combustion in fluidized bed). In: Raiko R, Kurki-Suonio I, Saastamoinen J, Hupa M, editors. Poltto ja palaminen. Jyväskylä, Finland: International Flame Research Foundation, Suomen kansallinen osasto; 1995.

[46] Spets J-P, A new multi-stage drying system. In: Proceeding of the 1st Nordic Drying Conference, Paper No. 13. Trondheim, Norway; June 27-29, 2001.

[47] Kunii D, Levenspiel O. Fluidization Engineering. Boston, London, Singapore, Sydney, Toronto, Wellington: Buttenworth-Heinemann; 1991.

[48] Brennan D. Process Industry Economics. United Kingdom: Institution of Chemical Engineers; 1998.

[49] Nord Pool. Elspot monthly prices. http://www.nordpool.com/nordpool/spot/index.html, February 28, 2006.

[50] The Finnish Ministry of Trade and Industry. Energiakatsaus (Energy Review). Helsinki, Finland: The Finnish Ministry of Trade and Industry, Energy Department; 4/2005.

[51] Menshutina NV, Kudra T. Computer Aided Drying Technologies. Drying Technology 2001; 19(8): 1825-1849.

[52] Jumah RY, Mujumdar AS. Control of Industrial Dryer. In: Mujumdar AS, editor. Handbook of Industrial Drying. New York: Marcel-Dekker; 1995. p. 1343-1367.

[53] The Finnish Bio Energy Society. Puupolttoaineiden laatuohje (Standard Instruction for Bio Fuel). Publications of Finbio. Finland: The Finnish Bio Energy Society; 1998.

[54] Pakowski Z, Mujumdar AS. Basic Process Calculations in Drying. In: Mujumdar AS, editor. Handbook of Industrial Drying. New York: Marcel Dekker; 1995. p. 71-111.

[55] Kotas TJ. The Exergy Method of Thermal Plant Analysis. Malabar, Florida: Krieger Publishing Company; 1995.

ISBN 978-951-22-8648-5
ISBN 978-951-22-8649-2 (PDF)
ISSN 1795-2239
ISSN 1795-4584 (PDF)