ADEME
Agence de l'Environnement et de la Maîtrise de l'Energie

KIT
Karlsruhe Institute of Technology

CITEPA

Liberté • Égalité • Fraternité
RÉPUBLIQUE FRANÇAISE
Ministère
de l'Écologie,
du Développement
durable
et de l'Énergie

TFTEI
TASK FORCE ON TECHNO ECONOMIC ISSUES

LRTAP
Long-range Transboundary Air Pollution

ESTIMATION OF COSTS OF REDUCTION TECHNIQUES FOR LCP

METHODOLOGY

TFTEI Technical Secretariat

28 May 2015

The methodology for estimation of costs of reduction techniques for LCP has been developed by the TFTEI Technical Secretariat. On this issue, developers are:

From KIT: Carmen Mayer, Ann-Kathrin Mueller, Simon Schulte Beerbuehl

From CITEPA: Nadine Allemand, Jean-Baptiste Vilmain

To assist the TFTEI Technical Secretariat to develop the methodology for cost estimations for LCP, a working group has been set up.

Participants were:

Jean-Guy Bartaire (TFTEI Co-chair); Richard Brandwood (E'on, United Kingdom); Nicolas Caraman (EDF, France); David Cooling (E'on, United Kingdom); Ivan Jankov (European Commission); Daniel Ladang (Total/CEFIC, Belgium); Hélène Lavray (Eurelectric, Belgium); Tiziano Pignatelli (TFTEI Co-chair); Arjan Plomp, (ECN, Netherlands); Koen Smekens (ECN, Netherlands); Jonathan Van der Kamp(EIFER, Germany).

Other experts contributed to delivering information to the TFTEI Technical Secretariat: the French group from the Chemical industry Union on coal boilers (chaired by Michel Monzain); Frans Van Aart (KEMA) in The Netherlands, manufacturers of abatement equipment (Hamon, Solvair, GE Air Filtration, CNIM), operators of combustion plants.

We apologise for not citing other experts who very kindly contributed to delivering information.

The TFTEI Technical Secretariat is grateful to all experts for helping us.

At its 33th session in December 2014, the Executive Body (EB decision 2014/2 http://www.unece.org/index.php?id=33291#/) of the United Nations Economic Commission for Europe (UN-ECE) Convention on Long Range Transboundary Air Pollution approved the upgrade of the Expert Group on Techno-Economic Issues (EGTEI) to Task Force on Techno-Economic Issues, TFTEI. This report often refers to EGTEI, but this designates TFTEI from now.

Table of Contents

Declaration of variables

Abbreviations	
ELV	Emission limit value
LHV	Lower Heating Value [MJ/kg]
LNB	Low NO_x burner technology
CLRTAP	Convention on Long range Transboundary Air Pollution
OFA	Overfire Air
SCR	Selective Catalytic Reduction
SNCR	Selective Non Catalytic Reduction
STP	Standard Temperature and Pressure (0°C, 1 atm)
PJFF	Pulse Jet Fabric Filter
ESP	ElectroStatic Precipitator
TSP	Total Suspended Particles
WAF	Water and Ash Free

Variables	
α	Biomass co-firing ratio
bs	Boiler size [MW]
C_i	Cost of substance i [EUR 2010 per unit]
c_i	Volumetric concentration of substance i [%-Vol.]
CAP	Capacity factor
$cons_i$	Mass consumption rate of substance i [mass per time]
f	Conversion factor
c/a	compressed air flow rate to actual air flow rate
$load_i$	Gravimetric load per volume of pollutant i [mg/Nm^3]
M_i	Molar mass of substance i [kg/kmol]
P	Electric Power [MW]
p	Interest rate [%]
ΔP or PD	Pressure drop
t	Time
x_i	Mass fraction of substance i [%-weight]
N	Standard conditions [273 K and 101 kPa]
V	Volume
η	Efficiency [%]
$\dot{m}_i$	mass flow rate of substance i [kg per time]
$\dot{v}_i$	Standard volumetric flow rate of substance i [Nm^3 per time]
$\dot{C}$	Cash flow rate
A	Area [m^2]
E	Electric field
A/C	Air to Cloth ratio (m/s)
MMD	Mass Mean Diameter [µm]
MMD_{rp}	Dust size change parameter

N	ESP field number
Pe	Dust penetration (%)
SCA	Specific Collecting Area (s/m)
T	Temperature [K]
WF	Working factor
ε_0	Free space permittivity (F/m)
N	Viscosity (kg/m/s)
Ω	Migration velocity [m/s]
NCR	Number of catalyst regenerations

Superindices

app	Application
bag	
baghouse	
BC	Back corona
bo	Boiler Outlet
Cage	
coal	
equip	Equipment, i. e. FDG, LNB, PJFF, SCR, SNCR
ESP	Electrostatic precipitator
ESP-mat	Material for ESP installation
fan	
PJFF	Pulse Jet Fabric filter
fuel	
gross	
inst	Installation
L	Loss
mat	Material
med	Media
MMD	Mass Mean Diameter
net	
PE	Polyethylene
RR	Rapping re-entrainment
SN	Sneakage

Subindices

adj	Adjusted
area	
ash	
avg	Average
bd	Sparking
c	Collection
cap	capital cost

cia	carbon in ash
Dry	dry conditions
dilution i	relative to the ith dilution
e	energy basis
ECP	Effective Collecting Plate
eq	Equivalent number of hours
elec	electricity basis
fan	
fix	Fix
full load	Full load power level
G	Gas
GC	Gross Cloth
i	C, H, O, N, S, SO_2, NO_x, H_2O, ash
in	Inlet
inv	investment
k	
m	mass basis
moist	Moisture
molar	molar basis
NC	Net Cloth
N-G	Net to Gross
O_2, act.	at actual O_2 concentration
O_2, ref.	at reference O_2 concentration
op	Operating
P	Penetrating
part load i	Part load power level i
r	Rapping
rep	Replacement
rib	retained in boiler-ratio [%]
s	Section
sec	Second
single	
spec	Specific
stoich	stoichiometric ratio
th	thermal basis
tot	Total
ut.el	utility electricity for fan
water	
wet	wet conditions
year	annual basis
λ	excess air ratio
comp	Compartment
air comp	compressed air

Preamble/Foreword

The first international legally binding instrument dealing with air pollution was the UNECE Convention on Long-range Transboundary Air Pollution (CLRTAP) adopted in 1979. This Convention supplied the framework for the First and Second Sulphur Protocol (1985, 1994), the Protocol on NO_x (1988) and on VOC (1991), with the objective of developing long-term policies to protect human health and the environment from the effects of air pollution. The most recent one is the Gothenburg Protocol from 1999, which came into force in May 2005 and has been amended in May 2012.

The amended Gothenburg Protocol (2012) aims at noticeably reducing acidification, eutrophication, tropospheric ozone formation and health impacts of fine particles ($PM_{2.5}$) by setting national emission reduction commitments (% reduction/2005 emissions) for the responsible pollutants, namely NO_x, SO_2, NMVOCs, NH_3 and $PM_{2.5}$, which have to be achieved by 2020. The Protocol further contains binding requirements in the form of emission limit values (ELVs) for both stationary and mobile sources, as well as fuel standards. These ELVs are presented in several technical annexes. A specific annex aims at reducing the emissions of ammonia from agricultural activities. Starting from the critical loads approach, and addressing several environmental problems and several pollutants simultaneously, this combined abatement strategy supplied a more cost-effective solution than treating pollutants or effects separately. It is why the Gothenburg Protocol is also called “Multi-pollutants and Multi-effects Protocol”.

Multinational strategies for the reduction of air pollution or greenhouse gases are mainly based on scenarios generated by means of Integrated Assessment Modelling (IAM). The new national emission reduction commitments in the amended Gothenburg Protocol were negotiated on the basis of results obtained by the GAINS (Greenhouse Gas and Air Pollution Interactions and Synergies) model, developed at the International Institute for Applied System Analysis (IIASA). GAINS is currently also being used for the revision of the NEC (National Emission Ceiling) Directive of 2001 by the European Commission.

The GAINS model estimates the internationally cost-optimal allocation of emission reductions. i.e. it determines where and how much emissions should be reduced to minimize the cost of removal and still meet pre-selected environmental targets (e.g. desired protection levels for vegetation, sensible ecosystems or human health) given by critical loads and levels, and constraints such as maximum allowable costs. Good knowledge of detailed technical and economic data for all relevant production processes and related abatement options is a crucial basis for credible IAM.

TFTEI established in December 2014 (previously EGTEI, Expert Group on Techno-Economic Issues established in 2001), is mandated by UNECE in the scope of the CLRTAP, to develop technical and economic data for relevant processes and related abatement techniques for stationary and mobiles sources.

The methodology for cost estimation of abatement options of SO_2, NO_x and TSP (Total Suspended Particulates) for Large Combustion Plants (LCP) with a thermal capacity of more than 50 MW_{th}, presented hereafter, aims at providing cost data for the following reduction techniques applied on large combustion plants using coal, heavy fuel oil and natural gas as well as biomass in co-combustion with coal.

Only boilers are considered (gas turbines and stationary engines could be examined later, in next steps if required).

Reduction techniques considered are as follows:

- NO_x: primary measures, SNCR (Selective Non Catalytic Reduction) and SCR (Selective Catalytic Reduction),
- TSP: electrostatic precipitator (ESP) and fabric filter (FF),
- SO_2: wet flue gas desulphurisation by limestone forced oxidation (LSFO – Limestone Forced Oxidation), semi dry (LSD - Lime Spray Dryer) and dry desulphurisation (DSI - Duct Sorbent Injection).

Costs are estimated for different regulatory objectives in term of ELVs (Emission Limit Values) assuming one boiler linked to a chimney.

To assist the TFTEI Technical Secretariat to develop the methodology for cost estimations for LCP, a working group has been set up.

Participants were:

Jean-Guy Bartaire (TFTEI Co-chair); Richard Brandwood (E'on, United Kingdom); Nicolas Caraman (EDF, France); David Cooling (E'on, United Kingdom); Ivan Jankov (European Commission); Daniel Ladang (Total/CEFIC, Belgium); Hélène Lavray (Eurelectric, Belgium); Tiziano Pignatelli (TFTEI Co-chair); Arjan PLOMP, ECN; Koen Smekens (ECN, Netherlands); Jonathan VAN DER KAMP(EIFER, Germany).

Other experts contributed to delivering information to the TFTEI Technical Secretariat: the French group from the Chemical industry Union on coal boilers (chaired by Michel Monzain); Frans Van Aart (KEMA) in The Netherlands, manufacturers of abatement equipment (Hamon, Solvair, GE Air Filtration, CNIM), operators of combustion plants.

We apologise for not citing other experts who very kindly contributed to delivering information.

1. Principles of the cost estimation

The methodology developed for estimating costs aims at being as much consistent and transparent as possible. To help with comparison of the data, cost components are clearly stated. As far as possible, recommendations of the Reference document on Economic and Cross Media Effects of the European Commission [EU Commission, 2006b] are taken into account.

For estimating costs, recent literature data and questionnaires have been used. Questionnaires have been developed and sent to LCP operators in order to collect recent investments and operating parameters to calculate operating costs. Questionnaires have been largely spread through the EGTEI working group members. Due to limited resubmission, statistical analyses could not be performed.

1.1. Composition of Costs

In the assessment process of BATs, the total annual costs, $\dot{C}_{tot}$ as well as the specific annual costs for abating the pollutant *i* are essential. They are defined according to equations 1-1 and 1-2:

$\dot{C}_{tot}\left[\frac{€}{year}\right] = \dot{C}_{cap}\left[\frac{€}{year}\right] + \dot{C}_{op}\left[\frac{€}{year}\right]$	1-1
$C_{tot,spec,i.}\left[\frac{€}{mass\ abated}\right] = \left(\dot{C}_{tot}\left[\frac{€}{year}\right]\right) \cdot \left(\dot{m}_{i,year}\right)^{-1}\left[\frac{year}{mass\ abated}\right]$	1-2

The total specific abatement costs per mass of pollutant *i*, $C_{tot,spec,i.}$ are calculated by dividing the total annual cost by the mass of abated pollutant $\dot{m}_{i,year}$, usually metric tons or kilograms. The specific total annual costs are calculated more thoroughly in the following chapters.

1.1.1. Investment

According to [EU Commission, 2006b], investments should include three components:

- Pollution control equipment expenditure,
- Installation expenditure,
- Contingency

Annex A presents the details of components which can be included in each category according to [EC 2006]. Literature data on investments very rarely give details on the components taken into account, so that comparisons are difficult. Investment for pollution control equipment and installation expenditure including permits, insurance, contingency etc. are usually given without taxes. To calculate the investment for retrofitting equipment to an existing combustion installation, a retrofit factor *r* shall represent the additional costs compared to an installation at a new plant.

For calculating costs of air pollution equipment at an annual level, the costs of the initial investment need to be spread onto each year of operation. The annualised capital cost can be calculated according to equation 1-3 with the parameters *p* (interest rate) and *n* (equipment technical or economic lifetime):

$\dot{C}_{cap}\left[\frac{€}{year}\right] = C_{inv} \cdot \frac{(1+p)^n}{(1+p)^n - 1} \cdot p$	1-3

In case of unknown life time of the control equipment, the lifetime is assumed to be equal to the lifetime of the power plant.

1.1.2. Operating Costs

Total operating costs are composed of fixed and variable operating costs, as shown in equation 1-4:

$\dot{C}_{op}\left[\frac{€}{year}\right] = \dot{C}_{op,fix}\left[\frac{€}{year}\right] + \dot{C}_{op,var}\left[\frac{€}{year}\right]$	1-4

The fixed operating costs $\dot{C}_{op,fix}$ are usually calculated as a percentage of the unit investment and include costs such as maintenance, insurance, wages[1], etc.

Variable operating costs $\dot{C}_{op,var}$ enclose costs for utilities such as electricity, waste disposal, reagents etc., as described in equation 1-5. The costs for disposal may be negative in case of the possibility of selling the residues (i.e. fly ash or gypsum):

$\dot{C}_{var}\left[\frac{€}{year}\right] = \sum \dot{C}^{unit}\left[\frac{€}{year}\right]$, unit ∈ {equipment, reagent, electricity, disposal}	1-5

1.2. Adaptation of temporal and currency differences

1.2.1. Adaptation of currency differences

Currency conversion to EURO from literature values in foreign currencies are done, if available at the reported conversion rates and stated explicitly. If no currency conversion rate was given, the yearly average of the conversion rate was determined and used for calculation.

1.2.2. Adaptation of temporal differences

Due to the time value of money, investment and costs cannot be compared without integrating the temporal aspect. To enable the comparison of costs or investments from different years, various indexes have been developed. One of these indexes, the Chemical Engineering Plant Cost Index (composite CEPCI)[2] shall be used in this document to allow for temporal adjustments (see

---

[1] It was the objective of the EGTEI technical secretariat to specify wage costs independently, when the revision of the cost methodology started. The working group decided to follow the common rules and include them in fixed operating costs finally due to lack of data.

[2] Published by Chemical Engineering Journal, www.che.com.

Table 1). The document works on EUR 2010:

Table 1: Cost elevation factors derived from CEPCI (www.che.com)

Year	2010	2009	2008	2007	2006	2005	2004	2003	2002	2001	2000
Multiple	1.00	1.06	0.96	1.05	1.10	1.18	1.24	1.37	1.39	1.40	1.40

1.3. Utility costs

Table 2 displays the default utility costs provided for calculating the operating costs of the pollutant abatement techniques:

Table 2: Default utility costs

Utility	Price	Unit
Electricity	30	€/MWh
Anhydrous ammonia	500	€/metric ton
Urea	300	€/metric ton
Limestone ($CaCO_3$) at a purity of 96%	20	€/metric ton
Lime (CaO) at a purity of 96 %	80	€/metric ton
Sulphur cost	70	€/metric ton

If they are available, country specific costs can be used otherwise.

The electricity consumption for overcoming the pressure drop by the abatement technologies will be calculated by using equation 1-6 [US EPA, 2002]:

$$cons_{elec}^{fan}\left[\frac{MWh}{year}\right] = 2.76 \cdot 10^{-8}\left[\frac{Pa \cdot MJ}{mbar \cdot s \cdot J}\right] \cdot \frac{\dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{year}\right] \cdot \Delta P[mbar]}{\eta_{fan}[\%]} \tag{1-6}$$

Fan efficiency ranges between 40% and 70% and is usually assumed to be 65% [US EPA, 2002].

2. Boiler Outlet Emission Loads

Combining the parametric set of boiler characteristics with the parametric set of fuel composition enables to calculate boiler outlet emission loads for all necessary pollutants. Hereby, the integration of information on boiler size and operating hours per year is provided, in order to derive yearly emission loads for the economic assessment.

Step 1 (subchapter Fuel consumption2.1) provides formulae for boiler capacity in terms of throughput.

Step 2 (subchapter 0) derives the boiler capacity factor.

Step 3 (subchapter 0) derives the volumetric boiler outlet emissions for given fuels.

In step 4 (0), formulae are given for integrating the effects of using different fuels within one boiler, for example the co-firing of low rank coals or biomass. According to the core idea of evaluating emission abatement techniques upon available information, calculations for boiler outlet emission loads will be provided on the basis of detailed, as well as, broad data.

2.1. Fuel consumption

The equation 2-1 links the thermal and electrical capacities of a power plant. It is important to notice, that thermal capacity (bs_{th}) is linked to the gross electrical output (bs_{el}^{gross}) via the gross electric efficiency (η^{gross}). Gross and net efficiency as well as boiler size are linked in equation 2-2 via the absolute respective relative plant electricity consumption, P^{plant}:

$bs_{el}^{gross}[MW_e] = bs_{th}[MW_{th}] \cdot \eta^{gross}\ [\%]$	2-1
$bs_{el}^{net}[MW_e] = bs_{el}^{gross}[MW_e] - P^{plant}[MW_e]$	2-2

The hourly full load fuel consumption $cons_e^{fuel}$can be derived via equation 2-3 on energy input basis and via eq. 2-4 on mass input basis via the lower heating value of the fuel:

$cons_e^{fuel}\left[\frac{GJ}{h}\right] = bs_{th}[MW_{th}]\ \cdot 3{,}600\ \left[\frac{s}{h}\right] \cdot 0.001\ \left[\frac{GJ}{MJ}\right]$	2-3
$cons_m^{fuel}\left[\frac{t}{h}\right] = \frac{cons_e^{fuel}\left[\frac{GJ}{h}\right]}{LHV^{fuel}\left[\frac{MJ}{kg}\right] \cdot \frac{1{,}000}{1{,}000}\left[\frac{GJ \cdot kg}{MJ\ \cdot t}\right]}$	2-4

For details on heating values of fuels and corresponding flue gas compositions, see chapter 0.

2.2. Capacity factor

The capacity factor allows deriving the annual electricity production and the catalyst lifetime. It is defined as the ratio of total annual full load hours to the total number of hours per year, as shown in equation 2-5:

$$CAP\,[\%] = \frac{t^{op}_{full\ load}\left[\frac{h}{y}\right]}{8{,}760\ \left[\frac{h}{y}\right]} \qquad \text{2-5}$$

The annual electric energy production is therefore derived as shown in equation 2-6:

$$W_{el,yearly}\left[\frac{MJ}{y}\right] = bs_{th,full\ load}[MW_{th}] * CAP\,[\%] * \eta^{gross}_{full\ load}[\%] * 31536000\left[\frac{s}{y}\right] \qquad \text{2-6}$$

Furthermore, the working factor can be calculated, which is defined as the ratio of total operating hours (independent of the load level) to the total number of hours per year, as shown in equation 2-7:

$$WF[\%] = \frac{t^{op}\left[\frac{h}{y}\right]}{8{,}760\ \left[\frac{h}{y}\right]} \qquad \text{2-7}$$

The equivalent number of hours worked at nominal capacity can be obtained either by a broad or by a detailed approach, depending on the availability of data. Those two approaches will be explained in the following.

2.2.1. Broad approach: full load consideration

The first approach is based on the direct input of a number of full load hours. The operating hours are expected to be equivalent, so only full load operation is taken into account in this approach. Depending on the operating scheme of the plant, this method might not be very accurate, however the calculation effort is low and there is not a lot of data necessary.

Table 3: Plant classification according to Strauß (2006) shows full-load operating hours for typical power plant classifications and can be used for orientation:

Table 3: Plant classification according to Strauß (2006)

Plant Classification	Full-load hours per year
Base load I	> 7,000 h/a
Base load II	5,000 - 7,000 h/a
Medium Load	2,000 - 5,000 h/a
Peak Load	< 2,000 h/a

2.2.2. Part load consideration

In practice, an LCP is rather likely not to be operated at full (nominal) load all the time. For example due to market fluctuations, the electrical – and therefore the thermal – energy output may be adapted to match the demand. Depending on the energy output, the gross electric efficiency changes as well.

In the detailed approach, the user can therefore consider part load operation and provide the data of the different load levels as shown in Table 4.

Table 4: Example of gross electric efficiency differences depending on the load level.

Load level	Number of hours per year t^{op}	Gross electric efficiency η^{gross}
500 MW (full load)	3000	39%
400 MW (80%)	1000	37%
350 MW (70%)	500	36%
300 MW (60%)	2000	35%
100 MW (40%)	500	34%

From the number of hours assigned to each energy level and by taking the change of gross electrical efficiency into account, equation 2-8 allows to calculate an equivalent number of full load hours.

$$t_{op,eq,full\ load}\left[\frac{h}{y}\right] = \sum_{i=load\ level} t_{load\ level\ i}^{op}\left[\frac{h}{y}\right] * \frac{\eta_{full\ load}^{gross}}{\eta_{load\ level\ i}^{gross}} * \frac{bs_{th,load\ level\ i}[MW_{th}]}{bs_{th,full\ load}[MW_{th}]} \tag{2-8}$$

This value can be used to determine the capacity factor, using equation 2-5.

2.3. Boiler and Fuel Characteristics

The aim of this subchapter is to derive the specific flue gas composition at boiler outlet on the basis of a given fuel specification. Depending on the detail of information about fuels used, a detailed and a rather general approach can be used. Solid and liquid fuels use a common methodology (subchapters 2.3.1), whereas for gases, a somewhat different approach has to be applied (subchapter 2.3.2). Solid and liquid fuels are usually characterised by mass analyses, i. e. elementary shares of carbon, hydrogen, sulphur, etc., while natural gases are usually characterised by component volume fractions (CH_4, C_2H_6, etc.).

At first, certain parameters, which characterise the combustion process, need to be specified. These operating parameters depend mainly on construction characteristics and plant operation. These factors are the excess air ratio (λ), the carbon-in-ash content (x_{cia}) and the fraction of ash retained in the boiler, i. e. the ratio of bottom ash to total ash. The ash-related parameters will be discussed and handled in subchapter 2.3.1.4. Combustion of natural gas usually emits no particulate matter, therefore TSP emission can be neglected. For excess air ratios, Strauß

[Strauß 2006] provides a range of typical ratios according to fuel and combustion technology for coal:

Table 5: Literature values for excess air ratios (Strauß, 2006)

Fuel / Technology	Typical values for λ
Oil	1.03-1.15
Gas	1.05-1.10
Coal	
dry bottom	1.2-1.3
wet bottom	1.15-1.25
grate	1.3-1.4

2.3.1. Detailed approach for solid and liquid fuels

Basis of this detailed approach is a fuel composition from elementary analysis of the water and ash free (*waf*) fuel, providing mass fractions of the relevant elementary components carbon (C), hydrogen (H), oxygen (O), nitrogen (N) and sulphur (S)[3] (CHONS). Assuming complete combustion, specific flue gas volumes as well as the lower heating value can be calculated from these CHONS data by mass balancing. Hereby, the following assumptions are made:

- Full oxidation of carbon to CO_2, no existence of CO or elementary carbon
- No nitrogen oxidation to nitrogen oxides
- Full oxidation of sulphur to SO_2, no existence of SO_3 and higher sulphur oxides or elementary sulphur

Large combustion plants usually emit CO, SO_3 and nitrogen oxides at mg/Nm³ levels – the error introduced by these assumptions is however rather small. Nevertheless, if values for NO_x, CO and SO_3 exist or are assumed, adjustment calculations can be added. Adjustments for elementary carbon can be meaningful, especially for PM emission load analysis and will therefore be performed in subchapter 2.3.1.2.

Table 6 shows some exemplary CHONS data for various hard coals from online literature surveys.

Table 7 shows equivalent data for typical liquid fuels used in large combustion plants. The exact ratios of the oils depend upon source and preparation, as these combustion fuels are not usually used in its raw state, but rather as a product of refining. In addition, HHV (Higher heating value) or LHV (Lower Heating Value) are well known characteristic figures and therefore, do not need to be calculated, as done for coal in equation 2-9.

---

[3] As this composition is water- and ash-free, the CHONS mass fractions should add up to 100%.

Table 6: Exemplary data for some important hard coals used in the LCP sector

Coal Mine	Country	Elementary composition (waf)					Ash	Moist.
		C	H	O	N	S		
Cerrejon	Columbia	83.4	4.95	9.47	1.37	0.81	8.41	11.83
Middelburg	South Africa	82.44	5.02	10.43	1.38	0.73	13.55	7.42
APC	Australia	88.58	4.73	4.22	1.46	1.00	11.12	10.27
Bachatsky	Russia	87.03	4.66	5.36	2.58	0.37	9.52	10.22
Bailey	USA	84.35	5.58	6.05	2.09	1.74	7.00	7.00
Blackwater	Australia	86.48	4.93	5.71	1.95	0.93	14.16	8.79
Douglas	South Africa	83.30	5.11	9.47	1.42	0.70	13.75	7.65
Elandsfontein	South Africa	88.16	4.86	4.91	1.43	0.64	12.74	9.00
Kleinkopje	South Africa	85.02	4.74	7.33	2.19	0.72	14.49	7.71
Kromdraai	South Africa	81.85	5.03	10.81	1.36	0.95	13.36	7.79

Table 7: Exemplary data for some important types of liquid fuels

Liquid fuel type	HHV [MJ/kg]	Elementary composition (waf)					Ash	Moist.
		C	H	O	N	S[4]		
Crude Oil	n/a	83-87	10-14	0.05-1,5	0.1-2	0.05-6	<1	<0.1
Gasoline	45.7	87	13	n/a	n/a	n/a	n/a	n/a
Diesel	47.0	84-86	13-15	n/a	n/a	<0.02	n/a	n/a
Biodiesel	40.0	77	12	11	n/a	0.01	n/a	n/a
Heavy Fuel Oil	43.0	86-88	8-10	n/a	n/a	1-5	0.50	0.1

---

[4] Commercial fuel oil, especially heavy fuel oil sulphur content varies strongly, as it is determined by refinery operations. Typically, heavy fuel oil can be separated into low sulphur (<0.5%), medium (0.5-2%) and high sulphur (>2%-vol.) heavy fuel oil.

2.3.1.1. Calculation of fuel lower heating value

The lower heating value of a fuel LHV^{fuel} can be derived upon its CHONS-characteristics according to equation 2-9 [Strauß 2006]:

$LHV^{fuel}\left[\frac{MJ}{kg}\right] = 33.9 \cdot x_C + 117.2 \cdot \left(x_H - \frac{x_O}{8}\right) + 10.5 \cdot x_S - 2.44 \cdot x_{moisture}$	2-9

2.3.1.2. Calculation of carbon-in-ash effect onto flue gas volume rates

The carbon-in-ash fraction x_{cia} represents the part of total carbon input, which will not be oxidised and therefore not turned into flue gases. Therefore it has to be subtracted from the carbon mass fraction, which will be used in the calculation of specific flue gas volumes. In general, the carbon-in-ash content varies and depends on fuel as well as on combustion characteristics. Standards for fly-ash usage in the construction industry (mainly road industry) limit the carbon-in-ash content to a maximum of 7% weight. Above this limit, fly-ash cannot be sold to this industry and needs to be land filled at a high cost. Therefore, power plants usually operate below 7%, mostly between 2% and 5% [EGTEI 2013]. Equations 2-10 and 2-11 show the necessary adjustments made to carbon ($x_{C,adj}$) and to ash (x_{ash+C}):

$x_{ash+C}\left[\frac{kg\ ash\ and\ carbon}{kg\ fuel}\right] = x_{ash}\left[\frac{kg\ ash}{kg\ fuel}\right] + \left(\frac{x_{ash}}{(1 - x_{cia})} - x_{ash}\right)\left[\frac{kg\ carbon}{kg\ fuel}\right]$	2-10
$x_{C,adj}\left[\frac{kg\ adj.carbon}{kg\ fuel}\right] = x_C\left[\frac{kg\ total\ carbon}{kg\ fuel}\right] - x_{cia}[\%] \cdot\ x_{ash+C}\left[\frac{kg\ ash\ and\ carbon}{kg\ fuel}\right]$	2-11

2.3.1.3. Calculation of ideal specific flue gas volume rates

Total flue gas volume consists of combustion products, nitrogen from combustion air and oxygen due to the use of excess air. For calculating the flue gas from the elementary CHONS-composition, Strauß [Strauß 2006] provides equation 2-12 for calculating the necessary combustion air, equation 2-13, and for calculating the specific flue gas volume flow at the excess air ratio λ, equation 2-14. Herein, x_i represents the element's mass fractions of the elementary analysis but corrected for ash and water[5]:

$v_{stoich,dry}^{flue\ gas}\left[\frac{Nm^3}{kg\ fuel}\right] = 8.899 \cdot x_{C,adj} + 20.96 \cdot\ x_H + 3.32 \cdot x_S + 0.80 \cdot x_N - 2.64 \cdot x_O$	2-12
$v_{stoich,dry}^{air}\left[\frac{Nm^3}{kg\ fuel}\right] = 8.899 \cdot x_{C,adj} + 26.514 \cdot\ x_H + 3.342 \cdot x_S - 3.340 \cdot x_O$	2-13

---

[5] Elementary analyses are usually provided free of water and ash (*waf*). For the following calculations, the mass fractions need to regard the water and ash masses, therefore x_i need to be corrected, if given at *waf*-level.

$v_{\lambda,dry}^{flue\ gas}\left[\frac{Nm^3}{kg\ fuel}\right] = v_{stoich,wet}^{flue\ gas}\left[\frac{Nm^3}{kg\ fuel}\right] + (1-\lambda)\cdot v_{stoich,dry}^{air}\left[\frac{Nm^3}{kg\ fuel}\right]$	2-14

The moisture content of the combustion air has been neglected, but could be integrated, if required by adjusting $v_{stoich,dry}^{air}$ in adding the moisture volume $v^{air\ moisture}$. The wet flue gas volume is important for calculating operating costs which are based on actual flue gas volume flow including moisture, such as for costs of pressure drop. Emission limit values (ELVs) on the other hand, are defined as emission loads (weight) per dry flue gas volume at standard conditions (0°C and 1 atm, resp. 273 K and 1.013 bar) and at a given reference oxygen concentration. Therefore, it is important to calculate the moisture content of the flue gas and to differentiate between the wet and dry flue gas volume flows $v_{\lambda,wet}^{flue\ gas}$ and $v_{\lambda,dry}^{flue\ gas}$, as detailed in equation 2-16.

The moisture volume $v^{moisture}$ can be derived from fuel moisture (given as mass fraction) $x_{moisture}$ and from fuel hydrogen content (from elementary analysis of the water and ash-free fuel) x_H in equation 2-15. The constant v_{molar} is the specific volume of 1,000 mol (1 kmol) of an ideal gas at standard temperature and pressure (1 atm, 0°C), 22.414 Nm³/kmol:

$\begin{aligned} v^{moisture}\left[\frac{Nm^3}{kg\ fuel}\right] &= n_{H_2O}\left[\frac{kmol}{kg\ fuel}\right]\cdot v_{molar}\left[\frac{Nm^3}{kmol}\right] \\ &= v_{molar}\left[\frac{Nm^3}{kmol}\right]\cdot\left(\frac{x_{moist}}{M_{H2O}}+\frac{x_H}{2*M_H}\right)\left[\frac{kmol}{kg\ fuel}\right] \end{aligned}$	2-15
$v_{\lambda,wet}^{flue\ gas} = v_{\lambda,dry}^{flue\ gas} + v^{moisture}$	2-16

2.3.1.4. Calculation of ash load

To calculate the ash load of the flue gas, it shall be assumed, that fuel ash will neither be gasified nor the oxidation states of the ash components changed (and thereby the mass in- or decreased). The ash either will leave the boiler as fly ash or as bottom ash. The ratio fly/bottom ash is depending on boiler and combustion characteristics and cannot be generalised. Therefore, either plant or literature data is needed. As seen in equation 2-17, the fly ash load can be derived by dividing the specific fly ash mass by the specific dry flue gas volume flow. As it is assumed, that the ash mass neither in- nor decreases, total fly ash is the difference between total ash input (from the fuel) and ash, which is retained in the boiler:

$$load_{ash,dry,O2ref}^{bo}\left[\frac{mg}{Nm^3}\right] = \frac{\left(x_{ash+C}\left[\frac{kg\ ash\ and\ carbon}{kg\ fuel}\right]\cdot(1-x_{ash,rib}\right)}{\dot{v}_{\lambda,dry}^{flue\ gas}\left[\frac{Nm^3}{kg\ fuel}\right]\cdot}\cdot f_{O2,corr}\cdot 10^6\left[\frac{mg}{kg}\right] \tag{2-17}$$

$$x_{ash,rib} = 1-\left(\frac{x_{fly\ ash}\left[\frac{kg}{kg\ fuel}\right]}{x_{total\ ash}\left[\frac{kg}{kg\ fuel}\right]}\right) \tag{2-18}$$

2.3.1.5. Calculation of SO_2 load

As mentioned, it shall be assumed that all sulphur shall be oxidized to SO_2. SO_3 loads are usually <5% of SO_2 emissions, therefore the calculation difference is small. Furthermore, sulphur oxide fractions adsorbed onto fly ash (mainly SO_3) are not taken into account, but could be added simply by adjusting total SO_2 and ash loads. With these simplifications, the SO_2 load can be calculated according to eq. 2-19:

$$load_{SO2,dry,O2ref}^{bo}\left[\frac{mg}{Nm^3}\right] = \frac{\frac{M_{SO2}\left[\frac{kg}{kmol}\right]}{Ms\left[\frac{kg}{kmol}\right]}\cdot x_S\left[\frac{kg\ S}{kg\ fuel}\right]\cdot\left[\frac{100-x_{ash\ [\%]}-x_{moist}[\%]}{10000}\right]\cdot\left[\frac{100-x_{S\ retainedinbo}[\%]}{100}\right]}{v_{\lambda,dry}^{flue\ gas}\left[\frac{Nm^3}{kg\ fuel}\right]}\cdot f_{O2,corr}\cdot 10^6\left[\frac{mg}{kg}\right] \tag{2-19}$$

2.3.1.6. Calculation of NO_x load

NO_x loads cannot be derived by mass balancing, as NO_x emissions will occur due to reaction of combustion air- and fuel-bound nitrogen with oxygen of the combustion air to form NO_x according to the mechanisms of thermal-, fuel- and prompt-NO_x formation. Contrary to SO_2 and ash, only a very small fraction of total nitrogen will be oxidized. Main determinants for NO_x generation are, according to literature, local temperature and concentration profiles (O_2-lean or O_2-rich regions). Therefore, no suitable way to estimate NO_x emissions without very detailed knowledge about a power plant or without measurement data does exist. Consequently, to estimate NO_x emissions for the purpose of this document, technique-related *averagely expected* NO_x emission loads, depending on fuel and combustion technology (low NO_x burner technology) shall be used. For the sake of this calculation procedure, NO_x loads at boiler outlet are assumed to be at the levels according to Table 8. Of course, if plant measurements or more precise results are known, they can be used instead. These values should reflect not only the installation of low NO_x burners themselves but rather the installation of a low NO_x combustion environment, i. e. including the installation of Overfire Air (OFA) or similar add-on technologies to burners.

Table 8: Reference NO_x boiler outlet emissions for solid fossil fuel combustion in boilers and process heaters according to LNB technology [mg/Nm³ ref. O_2] [EU Comission, 2006a; EU Comission, 2013]

		Hard coal / Bituminous coal		Lignite	
		without abatement	With low-NO_x-Burner	without abatement	with primary measures
Dry bottom boiler	horizontal firing system	1000 - 1500	500 - 650	400 - 700	200 - 500
	tangential firing system	600 - 900	400 - 650		
	vertical firing system	700 - 900			
Wet bottom boiler	cyclone firing system	1500 - 2500	1000 - 2000		

Table 9: Reference NO_x boiler outlet emissions for liquid fossil fuel combustion in boilers according to LNB technology in mg/Nm³ at 3%-vol. O_2 [EU COMMISSION, 2013]

	oil fired units
uncontrolled	800 - 1000
single primary measures	400 - 500
multiple primary measures	< 400

2.3.2. Detailed approach for gaseous fuels

Basis of this approach is a volumetric gas composition. Typical ingredients of natural gas found in the European gas grid are methane (CH_4) and light hydrocarbons (ethane, propane, butane), as well as nitrogen and carbon dioxide. Sometimes, small amounts of pentanes are ingredients as well. Oxygen and sulphur, as well as solid particles are no typical ingredients of grid-bound natural gas. Sulphuric compounds are used in mg/Nm³ loads for odorisation of natural gas.

Typical compositions of natural gas, which are available in Europe can be found in Table 10:

Table 10: Exemplary data for some important types of gaseous fuels

Natural Gas	HHV [MJ/kg]	Volumetric composition						
		CH_4	C_2H_6	C_3H_8	C_4H_{10}	CO_2	N_2	others[6]
Ekofisk, NO	44.42	86.08	8.02	2.66	0.86	1.61	0.52	0.25
Groningen, NL	35.17	81.6	2.78	0.37	0.13	1.00	14.0	0.12
Al Jubail, SA	48.15	71.1	25.6	1.50	0.20	1.00	0.60	0
Indonesia	40.99	83.19	3.36	2.65	1.43	8.35	0.34	0.68
Russia	39.09	95.74	1.00	0.07	0.02	1.30	1.80	0.07
United States	45.43	85.90	5.80	3.30	1.60	0.20	2.10	1.10
LNG	43.80	89.70	7.70	2.60	0	0	0	0

With this volumetric composition as a starting point, specific dry and wet flue gas volumes can be calculated from combustion statistics, assuming complete combustion and ideal gas laws (equations 2-20 and 2-21 for the specific, stoichiometric dry flue gas volume and for moisture). Combustion air demand can be calculated via stoichiometric oxygen demand (equation 2-22, assuming 21% O_2 concentration in air).Total flue gas volume at a predefined excess air ratio λ is a composition of these equations, either including or excluding the moisture (equation 2-23 for dry and 2-24 for the wet specific flue gas volume):

$$v_{stoich,dry}^{flue\ gas}\left[\frac{Nm^3}{Nm^3\ gas}\right] = 1\ Nm^3 * (c_{CH_4} + 2 * c_{C_2H_6} + 3 * c_{C_3H_8} + 4 * c_{C_4H_{10}} + c_{CO_2}) \tag{2-20}$$

$$v^{moisture}\left[\frac{Nm^3}{Nm^3\ gas}\right] = 1\ Nm^3 * \left(2 * c_{CH_4} + 3 * c_{C_2H_6} + 4 * c_{C_3H_8} + 5 * c_{C_4H_{10}}\right) \tag{2-21}$$

$$v_{stoich,dry}^{air}\left[\frac{Nm^3}{Nm^3\ gas}\right] = 1\ Nm^3 * \frac{(2 * c_{CH_4} + 3{,}5 * c_{C_2H_6} + 5 * c_{C_3H_8} + 6{,}5 * c_{C_4H_{10}})}{c_{O2\ in\ air}} \tag{2-22}$$

$$v_{\lambda,dry}^{flue\ gas} = v_{stoich,dry}^{flue\ gas} + (\lambda - 1) * v_{stoich,dry}^{air} \tag{2-23}$$

$$v_{\lambda,wet}^{flue\ gas} = v_{\lambda,dry}^{flue\ gas} + v^{moisture} \tag{2-24}$$

Higher heating value is typically supplied with the gas composition, as it is the billing unit. The lower heating value (LHV) can also be calculated according to the Vol.%-share of the gas components and their specific LHVs. Sulphur and particulate matter emissions are no concern of natural gas combustion and can be neglected.

---

[6] Higher hydrocarbons, treated as C_4H_{10}.

2.3.2.1. Calculation of NO_x emissions

For NO_x emissions, the same approach as for solid and liquid fuels has to be taken due to the same reasons. Table 11 provides some typical NO_x boiler outlet emissions, which can be found in literature.

Table 11: Reference NO_x boiler outlet emissions for g fossil fuel combustion in boilers according to LNB technology in mg/Nm³ at 3%-vol. O_2 [EU Commission, 2013]

	gas fired units
uncontrolled	150 - 400
single primary measures	75 - 150
multiple primary measures	< 75

2.3.3. General approach

The more general approach can be used when no information on the elementary fuel composition is known. Important input parameters are

- Specific flue gas volume (Nm³/kg fuel)
- Sulphur content of fuel (%-weight)
- Ash and moisture content of fuel (%-weight)

Usually, only the first parameter may be unknown, as the other parameters are essential for determining the coal quality. Strauß [Strauß 2006] provides a rough estimation of the flue gas factor and combustion air volume flow only using the LHV as input (equations 2-25 and 2-26). It is important to note, that equations 2-25 and 2-26 are only valid for coal. For heating oil and gas, equations 2-27 to 2-30 are given. The equations for natural gas require the gas to be expressed as weight input. Conversion from volumetric to weight input has to be done by dividing with the mass density of natural gas. If no specific density is known, general density values of 0.80 to 0.85 kg/Nm³ can be used instead. NO_x values have to be determined to the same way, as done for the detailed approach (subchapter 2.3.1.6).

$v_{stoich,wet}^{flue\ gas}\left[\frac{Nm^3}{kg\ coal}\right] = \left(0.2377 \cdot LHV^{coal} + 0.449\right)\left[\frac{Nm^3}{kg\ coal}\right] + \frac{x_{moist}}{M_{H2O}}\left[\frac{kmol}{kg}\right] \cdot \nu\ \left[\frac{Nm^3}{kmol}\right]$	2-25
$v_{stoich,dry}^{air}\left[\frac{Nm^3}{kg\ coal}\right] = \left(0.3163 \cdot LHV^{coal} + 0.566\right)\left[\frac{kg\ air}{kg\ coal}\right] \cdot \frac{1}{\delta_{air}}\left[\frac{Nm^3\ air}{kg\ air}\right]$	2-26
$v_{stoich,wet}^{flue\ gas}\left[\frac{Nm^3}{kg\ oil}\right] = \left(0.225 \cdot LHV^{oil} + 1.119\right)\left[\frac{Nm^3}{kg\ oil}\right] + \frac{x_{moist}}{M_{H2O}}\left[\frac{kmol}{kg}\right] \cdot \nu\ \left[\frac{Nm^3}{kmol}\right]$	2-27
$v_{stoich,dry}^{air}\left[\frac{Nm^3}{kg\ oil}\right] = \left(0.3437 \cdot LHV^{oil} - 0.425\right)\left[\frac{kg\ air}{kg\ oil}\right] \cdot \frac{1}{\delta_{air}}\left[\frac{Nm^3\ air}{kg\ air}\right]$	2-28

$$v^{flue\ gas}_{stoich,wet}\left[\frac{Nm^3}{kg\ gas}\right] = (0.2249 \cdot LHV^{gas} + 0.6476)\left[\frac{Nm^3}{kg\ gas}\right] + \frac{x_{moist}}{M_{H2O}}\left[\frac{kmol}{kg}\right] \cdot v\left[\frac{Nm^3}{kmol}\right] \tag{2-29}$$

$$v^{air}_{stoich,dry}\left[\frac{Nm^3}{kg\ gas}\right] = (0.3443 \cdot LHV^{gas} - 0.063)\left[\frac{kg\ air}{kg\ gas}\right] \cdot \frac{1}{\delta_{air}}\left[\frac{Nm^3\ air}{kg\ air}\right] \tag{2-30}$$

2.3.4. O_2 correction

Adjusting the oxygen concentration of flue gases is necessary, as emission limit values (ELV) are in general expressed at so called reference oxygen concentrations. Table 12 shows the O_2 reference concentrations, ELVs are expressed in the amended Gothenburg Protocol of the Convention on Long-Range Transboundary Air Pollution. It has to be noted that these reference concentrations may vary from legislation to legislation.

Table 12: Reference O_2 concentrations according to the amended Gothenburg Protocol

Fuel	Reference O_2 concentration
Solid fuels	6%-Vol.
Liquid fuels	3%-Vol.
Gaseous fuels in boilers and process heaters	3%-Vol.
Liquid and gaseous fuels in gas turbines	15%-Vol.

First, the actual oxygen concentration has to be calculated, as in equation 2-31. Second, the correction from actual to reference concentration can be done in equation 2-32.

$$c_{O2,act}[\%] = \frac{(1-\lambda) \cdot v_{air}\left[\frac{Nm^3}{kg\ fuel}\right]}{v^{flue\ gas}_{\lambda,dry}\left[\frac{Nm^3}{kg\ fuel}\right]} \tag{2-31}$$

$$f_{O2,corr} = \frac{(21 - c_{O2,act})}{(21 - c_{O2,ref})} \tag{2-32}$$

2.4. Fuel consumption, flue gas volume and NO$_x$ emission adjustment in case of part load consideration

The total annual flue gas volume is the sum of the flue gas volume at every load level taken into consideration:

$$\dot{v}_{\lambda,wet,year}^{flue\ gas} = \sum_{i=load\ level} \dot{v}_{\lambda,wet,year,load\ level\ i}^{flue\ gas} \tag{2-33}$$

For each load level, the fuel consumption per hour can be calculated with equation 2-34 and the annual fuel consumption for each level results of equation 2-35. Summing them up displays the total annual fuel consumption (equation 2-36).

$$\mathrm{cons}_{\mathrm{m,hour,load\ level\ i}}^{\mathrm{fuel}} \left[\frac{\mathrm{kg}}{\mathrm{h}}\right] = \frac{\mathrm{bs}_{\mathrm{th,load\ level\ i}}[\mathrm{MW_{th}}]}{\mathrm{LHV}^{\mathrm{fuel}}\left[\frac{\mathrm{MJ}}{\mathrm{kg}}\right]} * 3600 \left[\frac{\mathrm{s}}{\mathrm{h}}\right] \tag{2-34}$$

$$cons_{m,year,load\ level\ i}^{fuel} \left[\frac{kg}{y}\right] = cons_{m,hour,load\ level\ i}^{fuel} \left[\frac{kg}{h}\right] * t_{load\ level\ i}^{op} \left[\frac{h}{y}\right] \tag{2-35}$$

$$cons_{m,year}^{fuel} \left[\frac{kg}{y}\right] = \sum_{i=load\ level} cons_{m,year,load\ level\ i}^{fuel} \left[\frac{kg}{y}\right] \tag{2-36}$$

For each level, the dry and wet flue gas volumes per hour and per year can thus be calculated (equations 2-37 to 2-39).

$$\dot{v}_{\lambda,dry,hour,load\ level\ i}^{flue\ gas} \left[\frac{Nm^3}{h}\right] = cons_{m,hour,load\ level\ i}^{fuel} \left[\frac{kg}{h}\right] * v_{\lambda,dry}^{flue\ gas} \left[\frac{Nm^3}{kg}\right] \tag{2-37}$$

$$\dot{v}_{\lambda,wet,hour,load\ level\ i}^{flue\ gas} \left[\frac{Nm^3}{h}\right] = cons_{m,hour,load\ level\ i}^{fuel} \left[\frac{kg}{h}\right] * v_{\lambda,wet}^{flue\ gas} \left[\frac{Nm^3}{kg}\right] \tag{2-38}$$

$$\dot{v}_{\lambda,dry,year,load\ level\ i}^{flue\ gas} \left[\frac{Nm^3}{y}\right] = cons_{m,year,load\ level\ i}^{fuel} \left[\frac{kg}{y}\right] * v_{\lambda,dry}^{flue\ gas} \left[\frac{Nm^3}{kg}\right] \tag{2-39}$$

$$\dot{v}_{\lambda,wet,year,load\ level\ i}^{flue\ gas} \left[\frac{Nm^3}{y}\right] = cons_{m,year,load\ level\ i}^{fuel} \left[\frac{kg}{y}\right] * v_{\lambda,wet}^{flue\ gas} \left[\frac{Nm^3}{kg}\right] \tag{2-40}$$

Summing them up finally displays the total dry and wet flue gas volume per year.

$$\dot{v}_{\lambda,dry,year}^{flue\ gas} = \sum_{i=load\ level} \dot{v}_{\lambda,dry,year,load\ level\ i}^{flue\ gas} \tag{2-41}$$

The whole methodology to derive NO$_x$ emissions has been described above. Nevertheless, part load consideration may be taken in account. Indeed, the specific NO$_x$ emission (related to fuel consumption) is depending on the load level, due to the variation of the combustion temperature. Hence, it is possible to adjust the amount of NO$_x$ emissions at the different load levels based on an emission factor in relation to the full load emissions.

NO_x emissions per hour can be derived for every load level.

$$\dot{m}^{NOx}_{\lambda,dry,hour,load\ level\ i}\left[\frac{kg}{h}\right] = \frac{\dot{v}^{flue\ gas}_{\lambda,dry,hour,load\ level\ i}\left[\frac{Nm^3}{h}\right] * load^{NOx}_{dry,ref\ O_2,load\ level\ i}\left[\frac{mg}{Nm^3}\right]}{1.000.000\left[\frac{mg}{kg}\right]} \tag{2-42}$$

Therefrom the annual amount of NO_x emissions for each load level can be calculated.

$$\dot{m}^{NOx}_{\lambda,dry,year,part\ load,i}\left[\frac{t}{y}\right] = \frac{\dot{v}^{flue\ gas}_{\lambda,dry,year,load\ level\ i}\left[\frac{Nm^3}{y}\right] * load^{NOx}_{dry,ref\ O_2,load\ level\ i}\left[\frac{mg}{Nm^3}\right]}{1.000.000.000\left[\frac{mg}{t}\right]} \tag{2-43}$$

And the volume of total emissions is derived by summing them up.

$$\dot{m}^{NOx}_{\lambda,dry,year}\left[\frac{t}{y}\right] = \sum_{i=load\ level} \dot{m}^{NOx}_{\lambda,dry,year,load\ level\ i}\left[\frac{t}{y}\right] \tag{2-44}$$

2.5. Integration of Biomass Co-firing

Biomass can be co-fired via three different methods: direct, indirect and parallel co-firing. In this methodology only direct co-combustion is considered. This assumption needs to be made in order to be able to analyse the effect of biomass co-combustion on a given unit, which may co-combust biomass or may alternatively fire purely fossil fuels. Both other types of co-combustion would require modifications in fuel preparation and injection into the furnace, so that substantial investments need to be made. Typical co-firing ratios using the method of direct co-firing reach up until 20% on a mass-basis [ECN 2013].

Regarding the type of biomass only wood is included. Co-firing of straw and other biomass types is less common than of wood and requires more cautioness with regard to additional flue gas components (chlorine, fluorine and higher contents of alkaline metals) and their effects onto the combustion, the ash and gypsum as well as SCR deactivation. In Table 13 the elemental composition of different exemplary wood types are given.

Table 13: Exemplary elemental composition of different wood types [Bunbury, 1925]

Wood type	Composition of exemplary biomass (water and ash free, waf)			
	C	H	O	N
Oak	50,64	6,23	41,85	1,28
Beech	50,89	6,07	42,11	0,93
Birch	50,61	6,23	42,04	1,12
Pine	51,39	6,11	41,56	0,94
Spruce	51,39	6,11	41,56	0,94

Effect of co-firing on boiler outlet emissions:

The effect of firing up to 20% biomass (mass-basis) onto the boiler outlet emissions will be discussed in the following. The following variables are subject to changes:

- Specific flue gas volume per kilogram of fuel
- SO_2 emission load
- TSP emission load
- NO_x emission load

The first three effects can be integrated using already described formulae for coal combustion, as biomass features the same ingredients as coal. For calculating the emission load $\dot{m}_{i,annual}$ from biomass co-firing, first the emissions for both pure coal and pure biomass combustion are calculated according to eq. 2-45 and 2-46 for each of the fuels. The results of eq. 2-45 and 2-46 would therefore assume pure combustion.

$\dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{y}\right] = v_{\lambda,dry}^{flue\ gas}\left[\frac{Nm^3}{kg\ fuel}\right]\cdot cons_m^{fuel}\left[\frac{kg\ fuel}{h}\right]\cdot t_{op}^{full\ load}\left[\frac{h}{y}\right]$	2-45
$\dot{m}_{i,annual}\left[\frac{t}{y}\right] = load_{i,dry}\left[\frac{mg}{Nm^3}\right]\cdot \dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{y}\right]\cdot 10^{-9}\left[\frac{t}{mg}\right]$	2-46

To derive the co-firing concentration, at first the total annual flue gas volume according to the co-combustion ratio need to be calculated (eq. 2-47). Therefore, the values of the biomass are multiplied with the biomass co-firing ratio α and the coal with (1- α). The pollutant load can be derived now by simply dividing eq. 2-48 with eq. 2-47.

$\dot{v}_{\lambda,dry,year,total}^{flue\ gas}\left[\frac{Nm^3}{y}\right] = \dot{v}_{\lambda,dry,year,coal}^{flue\ gas}\left[\frac{Nm^3}{y}\right]\cdot(1-\alpha) + \dot{v}_{\lambda,dry,year,biomass}^{flue\ gas}\left[\frac{Nm^3}{y}\right]\cdot\alpha$	2-47
$\dot{m}_{i,annual}\left[\frac{t}{y}\right] = \dot{m}_{i,annual,coal}\left[\frac{t}{y}\right]\cdot(1-\alpha) + \dot{m}_{i,annual,biomass}\left[\frac{t}{y}\right]\cdot\alpha$	2-48

It is obvious, that SO_2 emissions will decrease with increasing co-firing share due to the lesser sulphur concentration of the fuel. The effect on TSP emissions cannot be foreseen generally, but in most cases, boiler outlet emissions are expected to rise slightly.

The effect of biomass co-firing on NO_x emissions is highly site-specific, as NO_x emissions are mainly determined by the combustion characteristics and combustion technology. According to the LCP BREF, NO_x emissions are expected to decrease due to the smaller nitrogen content of the fuel [EU Commission, 2006a]. American combustion measurements have shown that NO_x emissions decrease due to less production of thermal NO_x when co-firing at old boilers with boiler outlet emissions of more than 1,000 mg/Nm³. However, no effect on NO_X emissions as well as no change in efficiency has been observed at measurements in newly erected coal-fired power plants and at co-firing ratios of less than 20% [ECN 2013].

3. Evaluation of NO_x abatement techniques

For abating NO_X emissions primary and secondary measures can be applied. Primary measures aim at reducing the amount of NO_x formed during combustion by optimising boiler or turbine parameters with respect to emissions. Secondary measures include reagents that reduce NO_x. In the following paragraphs, the operating costs for NO_x abatement techniques are calculated.

3.1. Primary Measures

As primary measure for reducing NO_x emissions low NO_x Burners (LNB) are defined to be state of the art. These are designed to control air mixing at each burner in order to create larger and more branched flames and therefore to reduce NO_x formation. Plants without LNBs under IPPC are rather rare. Therefore, only LNBs are considered in this methodology.

To calculate the total costs per year of primary measures and the specific NO_x reduction costs the following equations can be used.

$$C_{inv,LNB}[€] = C_{inv,LNB,spec}\left[\frac{€}{kW_{th}}\right] \cdot bs_{th}[kW_{th}] \tag{3-1}$$

$$\dot{C}_{cap,LNB}\left[\frac{€}{y}\right] = C_{inv,LNB} \cdot \frac{(1+p)^n}{(1+p)^n - 1} \cdot p \tag{3-2}$$

$$\dot{C}_{op,fix,LNB}\left[\frac{€}{y}\right] = C_{inv,LNB}[€] \cdot OM_{fix}\left[\frac{\%}{y}\right] \tag{3-3}$$

The investment for a LNB is given by the specific investment per kW_{th} of the LCP (see Annex A) multiplied with the installed thermal capacity. The fixed operating costs are calculated with the factor for fixed operating and management costs (see equation no. 3-3). According to equation 1-3 the capital costs for LNB can be calculated as in equation no.3-2 by multiplying investment and interest rate. As the variable operating costs are zero, total costs for the LNB are calculated according to eq. 3-4.

$$\dot{C}_{tot,LNB}\left[\frac{€}{y}\right] = \dot{C}_{cap,LNB}\left[\frac{€}{y}\right] + \dot{C}_{op,fix,LNB}\left[\frac{€}{y}\right] \tag{3-4}$$

The specific NO_X abatement cost can then be calculated by dividing the total cost for LNB by the total NO_X emissions abated from upgrading primary measures.

3.2. Secondary Measures

The most common secondary measures are Selective Catalytic Reduction (SCR) and Selective Non Catalytic Reduction (SNCR). Secondary measures abate NO_x by reducing NO and NO_2 to pure N_2 and water (plus CO_2 in the case of urea) with the use of appropriate reagents such as ammonia or urea.

3.2.1. Selective Catalytic Reduction

SCRs are mostly built up of multiple layers and use a catalyst to improve the abatement efficiency. In the following subchapters the operating properties are described and included in the calculation of the operating costs of SCRs.

3.2.1.1. Pressure drop calculation

A SCR is built up of multiple layers, which influence the pressure drop significantly. Additionally to the pressure drop of each layer, the pressure drop of injection and mixing as well for the casing itself contribute to the necessary electricity and, therefore, the electricity cost. The total pressure drop of an SCR can be calculated by summarizing the above mentioned influencing factors.

$$PD_{SCR}[mbar] = PD_{inj}[mbar] + PD_{SCR,ex}[mbar] + PD_{layer}[mbar] \cdot n_{layer} \tag{3-5}$$

The calculation of the electricity costs which occur to overcome the pressure drop is calculated in below.

3.2.1.2. Catalyst cost

The costs for the catalyst are dependent on the catalyst volume, catalyst lifetime and catalyst management strategy (number of and interval between catalyst regeneration).

For these calculations a reference SCR unit may have three active and one spare layer. These can be regenerated after a certain time which leads to a number of possible catalyst regenerations. If no regeneration of the catalyst is done then the lifetime is mostly reduced. The choice if the catalyst is regenerated or not lies within the responsibility of the plant operator.

$$V_{total}^{catalyst}[m^3] = V_{spec}^{catalyst}\left[\frac{m^3}{MW_{th}}\right] \cdot bs_{th}[MW_{th}] \tag{3-6}$$

$$T_{lifetime,op}^{catalyst}[y] = \frac{T_{lifetime}^{catalyst}[h] * (NCR + 1)}{CAP \cdot 8{,}760\left[\frac{h}{y}\right]} \tag{3-7}$$

$$\dot{C}_{catalyst}\left[\frac{€}{y}\right] = \frac{V_{total}^{catalyst}[m^3] \cdot c_{catalyst}\left[\frac{€}{m^3}\right] + NCR \cdot V_{total}^{catalyst}[m^3] \cdot c_{catalyst,reg}\left[\frac{€}{m^3}\right]}{T_{lifetime,op}^{catalyst}[y]} \tag{3-8}$$

Data from literature, previous EGTEI information and LCP operators are given in Table 14 as a benchmark. The lifetime is strongly dependent on the catalyst management strategy and the fuel characteristics (heavy metals, biomass etc.).

Table 14: Catalyst Cost and Design Data from Questionnaires (example for solid fuel plants)

Catalyst Cost & Design Data for Hard Coal Fired Units				
	Min	Max	Unit	No. of references
Spec. cat. Requirement	0.5	0.59	m^3/MW_{th}	2
Total cat. Lifetime	24,000	36,000	h	2
No. of cat. regenerations	0	3		3
Spec. cat. cost	4,000	5,000	€/m³	2
Spec. cat. regeneration cost	2,500		€/m³	1

3.2.1.3. Reagent consumption

For reducing NO_x molecules, corresponding reducing reagents are necessary. These are mostly ammonia or urea but also other reagents can be used. Depending on the chosen reagent the stoichiometric factors and necessary reagent amount are determined.

Their amount can be calculated by stoichiometric factors which have been taken from literature and questionnaire feedback and have a range of 0.8 to 0.9 for SCR. The corresponding equations are given in eq. 3-9 to eq. 3-10. It is important to note, that the NO_x concentrations in eq. 3-9 need to be at actual oxygen concentration, and not at corrected oxygen concentration. This is due to the fact, that the annual flue gas volumes relate to actual oxygen concentration as well and would need to be adjusted to the corrected oxygen concentration otherwise. Consequently, if the SCR shall ensure a specific ELV, expressed at reference oxygen concentration, this ELV must be divided by the oxygen correction factor (eq. 2-22) to derive $c_{NOx,load\ emit}$ at the actual oxygen concentration.

$\dot{m}_{NOx,abated,SCR}\left[\frac{t}{y}\right] = \left(c_{NOx,in}\left[\frac{mg}{Nm^3}\right] - c_{NOx,load\ emit}\left[\frac{mg}{Nm^3}\right]\right)\cdot \dot{v}^{flue\ gas}_{\lambda,dry,year}\left[\frac{Nm^3}{y}\right]\cdot 10^{-9}\left[\frac{t}{mg}\right]$	3-9
$\dot{m}_{reagent\ i,SCR}\left[\frac{t}{y}\right] = \frac{M_{reagent\ i}}{M_{NOx}}\cdot \dot{m}_{NOx,abated,SCR}\left[\frac{t}{y}\right]\cdot SR_{SCR}$	3-10

Reagents are usually not bought pure, but as an aqueous solution, mixed with water at a certain concentration $a_{dilution\ 1,reagent\ i}$. As an example, urea is applied as a 50% solution, while ammonia is 29.4% diluted in [US EPA, 2002].

The reagents are bought, delivered and stored at this specific concentration. Before they are injected into the system, they might be further diluted with water to reach a lower dilution concentration $a_{dilution\ 2,reagent\ i}$. The following equations calculate the mass flow of first-dilution reagent (i.e. at the state of delivery) depending on the mass flow of the NO_x that needs to be abated. Another result is the mass flow of water that has to be added to first-dilution reagent in order to obtain the second-dilution concentration (at the state of injection).

$$\dot{m}_{reagent\ i,SCR,dilution\ 1}\left[\frac{t}{y}\right] = \frac{1}{a_{reagent\ i,SCR,dilution\ 1}} \cdot \dot{m}_{reagent\ i,SCR,SCR}\left[\frac{t}{y}\right] \tag{3-11}$$

$$\dot{m}_{water,SCR,dilution\ 2}\left[\frac{t}{y}\right] = \left(\frac{1}{a_{reagent\ i,SCR,dilution\ 2}} - \frac{1}{a_{reagent\ i,SCR,dilution\ 1}}\right) \cdot \dot{m}_{reagent\ i,SCR}\left[\frac{t}{y}\right] \tag{3-12}$$

As the prices of the compounds are input parameters, the final reagent cost can be calculated as shown below:

$$\dot{C}_{reagent\ i,SCR}\left[\frac{€}{y}\right] = \dot{m}_{reagent\ i,SCR,dilution\ 1}\left[\frac{t}{y}\right] \cdot C_{spec,reagent\ i,dilution\ 1}\left[\frac{€}{t}\right] + \dot{m}_{water,SCR,dilution\ 2}\left[\frac{t}{y}\right] \cdot C_{spec,water}\left[\frac{€}{t}\right] \tag{3-13}$$

3.2.1.4. Additional power consumption

Power consumption is needed for reagent storage, distribution and injection as well as for the compensation of the corresponding pressure drop. It is assumed, that the electricity consumption has no linear dependence of the load level, but is only slightly lower in case of part load operation. Therefore the work factor (WF) is applied in this case instead of the capacity factor (CAP) to avoid an underestimation of the electricity demand.

$$\dot{C}_{electricity,SCR}\left[\frac{€}{y}\right] = \left(cons_{utility,SCR}\left[\frac{MWh}{h}\right] \cdot WF \cdot 8{,}760\left[\frac{h}{y}\right] + cons_{pd,SCR}\left[\frac{MWh}{Mio.\ Nm^3 \cdot mbar}\right] \cdot PD_{SCR}[mbar] \cdot \frac{\dot{v}_{wet}^{flue\ gas}\left[\frac{Nm^3}{y}\right]}{10^6}\right) \cdot C_{electricity,spec.,SCR}\left[\frac{€}{MWh}\right] \tag{3-14}$$

According to equation 1-4 the operating costs can be calculated as stated below (equation 3-15 to 3-19).

$C_{inv,SCR}[€] = C_{inv,SCR,spec}\left[\frac{€}{MW_{th}}\right] \cdot bs_{th,full\ load}[MW_{th}]$	3-15
$\dot{C}_{cap,SCR}\left[\frac{€}{y}\right] = C_{inv,SCR}[€] \cdot \frac{(1+p)^n}{(1+p)^n - 1} \cdot p$	3-16
$\dot{C}_{op,SCR}\left[\frac{€}{y}\right] = \dot{C}_{op,fix,SCR}\left[\frac{€}{y}\right] + \dot{C}_{op,var,SCR}\left[\frac{€}{y}\right]$	3-17
$\dot{C}_{op,fix,SCR}\left[\frac{€}{y}\right] = C_{inv,SCR}[€] \cdot OM_{fix}\left[\frac{\%}{y}\right]$	3-18
$\dot{C}_{op,var,SCR}\left[\frac{€}{y}\right] = \dot{C}_{reagent,SCR}\left[\frac{€}{y}\right] + \dot{C}_{electricity,SCR}\left[\frac{€}{y}\right] + \dot{C}_{Catalyst}\left[\frac{€}{y}\right]$	3-19

3.2.1.5. Effect of biomass co-firing on SCR

In case of biomass, co-firing biomass can reduce the lifetime of the catalyst by poisoning it. The deactivation of the catalyst is influenced by phosphorus and silica compounds [Koppejan, 2010]. According to Jensen-Holm et al., a co-firing rate of less than 20% does not have any noticeable impact on catalyst lifetime. State of the art is rather 10% of biomass share and not more than 20%. Therefore, only low co-firing ratios are regarded in this methodology. If the co-firing ratio exceeds 20% there might be an effect on the catalyst. Reducing the lifetime of the catalyst leads to higher SCR costs due to a larger amount of needed catalyst (see Harding).

3.2.2. Selective Non Catalytic Reduction

Due to economic aspects and ELV requirements the application of SNCR might be limited to plants below 500 MW_{th} for application with high abatement efficiency. SNCR tests show that with rising plant sizes the reduction efficiencies decrease due to injection and mixing constraints. Table 15 provides values for maximal reduction efficiencies of SNCR. Maximum achievable reduction rates decrease for larger capacities due to increasing reagent and flue gas mixing challenges, which may lead to rising ammonia slip.

This table provides achieved values from SNCR retrofits. The case specific maximum achievable NO_x abatement efficiency varies from plant to plant and is in general dependent upon the operating temperature (determined by the flue gas temperature at the point of reagent injection and mixing), the stoichiometric reagent-to-NO_x ratio, and of course, the degree of mixing. A general interdependence between the temperature, stoichiometric ratio of ammonia, resp. urea, and NO_x to the NO_x abatement efficiency can be seen in Annex E. When analysing these diagrams, it has to be borne in mind, that ammonia ELVs (due to ammonia slip) will limit the stoichiometric ratio in most cases and thereby inhibit abatement efficiencies of more than the ones displayed in Table 15.

Table 15: Literature Data for SNCR Efficiencies

SNCR Efficiency	
Maximum Achievable SNCR Reduction Rates	
Plant Size	Max. Reduction
< 100 MW_{th}	60%
100 - 300 MW_{th}	55%
300 - 500 MW_{th}	47.5%
500 - 700 MW_{th}	40%
> 700 MW_{th}	35%

3.2.2.1. Reagent Consumption

The amount of reagents can be calculated by stoichiometric factors which are derived from literature and questionnaire feedback and have a range of 1.5 to 2.0 for SNCR. It is important to note, that the NO_x concentrations in eq. 3-20 need to be at actual oxygen concentration, and not at corrected oxygen concentration. This is due to the fact, that the annual flue gas volumes relate to actual oxygen concentration as well and would need to be adjusted to the corrected oxygen concentration otherwise. Consequently, if the SNCR shall ensure a specific ELV, expressed at reference oxygen concentration, this ELV must be divided by the oxygen correction factor (eq. 2-22) to derive $c_{NOx,load\ emit}$ at the actual oxygen concentration.

$$\dot{m}_{NOx,abated,SNCR}\left[\frac{t}{y}\right] = \left(c_{NOx,in}\left[\frac{mg}{Nm^3}\right] - c_{NOx,load\ emit}\left[\frac{mg}{Nm^3}\right]\right)\cdot \dot{v}^{flue\ gas}_{\lambda,dry,year}\left[\frac{Nm^3}{y}\right]\cdot 10^{-9}\left[\frac{t}{mg}\right] \tag{3-20}$$

$$\dot{m}_{reagent\ i,SNCR}\left[\frac{t}{y}\right] = \frac{M_{reagent\ i}}{M_{NOx}}\cdot \dot{m}_{NOx,abated,SNCR}\left[\frac{t}{y}\right]\cdot SR_{SNCR} \tag{3-21}$$

Reagents are usually not bought pure, but as an aqueous solution, mixed with water at a certain concentration $a_{dilution\ 1,reagent\ i}$. As an example, urea is applied as a 50% solution, while ammonia is 29.4% diluted in [US EPA, 2002].

The reagents are bought, delivered and stored at this specific concentration. Before they are injected into the system, they might be further diluted with water to reach a lower dilution concentration $a_{dilution\ 2,reagent\ i}$. The following equations calculate the mass flow of first-dilution reagent (i.e. at the state of delivery) depending on the mass flow of the NO_x that needs to be abated. Another result is the mass flow of water that has to be added to first-dilution reagent in order to obtain the second-dilution concentration (at the state of injection).

$$\dot{m}_{reagent\ i,SNCR,dilution\ 1}\left[\frac{t}{y}\right]=\frac{1}{a_{reagent\ i,dilution\ 1}}\cdot\dot{m}_{reagent\ i,SNCR}\left[\frac{t}{y}\right] \tag{3-22}$$

$$\dot{m}_{water,SNCR,dilution\ 2}\left[\frac{t}{y}\right]=\left(\frac{1}{a_{reagent\ i,SNCR,dilution\ 2}}-\frac{1}{a_{reagent\ i,SNCR,dilution\ 1}}\right)\cdot\dot{m}_{reagent\ i,SNCR}\left[\frac{t}{y}\right] \tag{3-23}$$

As the prices of the compounds are input parameters, the final reagent cost can be calculated as shown below:

$$\dot{C}_{reagent\ i,SNCR}\left[\frac{€}{y}\right]=\dot{m}_{reagent\ i,SNCR,dilution\ 1}\left[\frac{t}{y}\right]\cdot C_{spec,reagent\ i}\left[\frac{€}{t}\right]+\dot{m}_{water,SNCR,dilution\ 2}\left[\frac{t}{y}\right]\cdot C_{spec,water}\left[\frac{€}{t}\right] \tag{3-24}$$

3.2.2.2. Pressure Drop

Contrary to the SCR the pressure drop of the SNCR is only dependent on the injection and mixing process. Therefore, the pressure drop for calculating the needed electricity is calculated as follows.

$$PD_{SNCR}=PD_{inj} \tag{3-25}$$

3.2.2.3. Additional power consumption

Power consumption is needed for reagent storage, distribution and injection as well as for the compensation of the corresponding pressure drop.

$$\dot{C}_{electricity,SNCR}\left[\frac{€}{y}\right]=\left(cons_{utility,SNCR}\left[\frac{MWh}{h}\right]\cdot WF\cdot 8{,}760\left[\frac{h}{y}\right]+cons_{pd,SNCR}\left[\frac{MWh}{Mio.\ Nm^3\cdot mbar}\right]\cdot PD_{SNCR}[mbar]\cdot\dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{y}\right]\cdot 10^{-6}\right)\cdot C_{electricity,spec.,SNCR}\left[\frac{€}{MWh}\right] \tag{3-26}$$

According to the explanation in chapter 3.2.1.4 here as well the work factor (WF) ist applied instead of the capacity factor (CAP).

3.2.2.4. Total costs

SNCR does not utilize catalysts. Therefore, the operating costs for SNCR only include cost for electricity and reagent. These are calculated according to the cost functions given in chapter 1.1 and respectively to the calculation of SCR. The specific NO_X abatement costs can be calculated by dividing the total cost of capital costs and operational costs by the total NO_X abated.

$C_{inv,SNCR}[€] = C_{inv,SNCR,spec}\left[\frac{€}{MW_{th}}\right] \cdot bs_{th,full\ load}[MW_{th}]$	3-27
$\dot{C}_{cap,SNCR}\left[\frac{€}{y}\right] = C_{inv,SNCR} \cdot \frac{(1+p)^n}{(1+p)^n-1} \cdot p$	3-28
$\dot{C}_{op,SNCR}\left[\frac{€}{y}\right] = \dot{C}_{op,fix,SNCR}\left[\frac{€}{y}\right] + \dot{C}_{op,var,SNCR}\left[\frac{€}{y}\right]$	3-29
$\dot{C}_{op,fix,SNCR}\left[\frac{€}{y}\right] = C_{inv,SNCR}[€] \cdot OM_{fix}\left[\frac{\%}{y}\right]$	3-30
$\dot{C}_{op,var,SNCR}\left[\frac{€}{y}\right] = \dot{C}_{reagent,SNCR}\left[\frac{€}{y}\right] + \dot{C}_{electricity,SNCR}\left[\frac{€}{y}\right]$	3-31

4. Evaluation of dust abatement techniques

4.1. Introduction

To control dust emission from Large Combustion Plant (LCP), two techniques are considered in this report: Pulse Jet Fabric Filter and Electrostatic Precipitator (PJFF and ESP).

US EPA *Air Pollution Cost Control Manual* [EPA, 2002] is a reference in bottom-up approach for cleaning process cost estimations. PJFF and ESP investment calculations developed in this working document are based on the US EPA general approach and adapted with current data and reference literature.

4.2. Fabric filter

To scale-up a PJFF unit, US EPA methodology develops cost equations based on total filtration area. This main design parameter is determining through volumetric flow rate and Air to Cloth ratio. Figure 4-1 presents the logical diagram for total filtration area determination with involving equations and Ref.boxes.

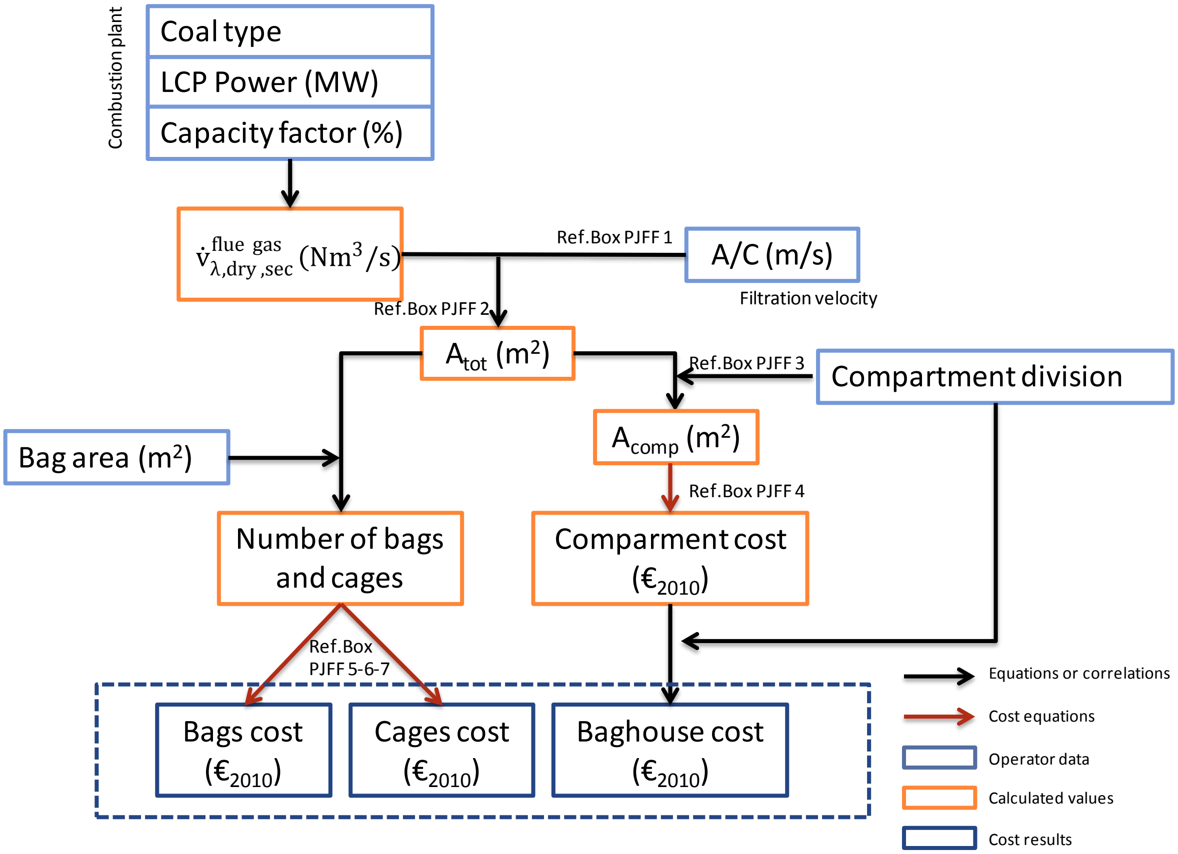


Figure 4-1-Logical diagram for PJFF cost determination

4.2.1. Air-to-Cloth ratio and filtration Area

Filtration theory relies on particles penetration potential through a filtering layer. To scale up a PJFF unit, a ratio between the volumetric gas flow and the filtration area is used. This ratio is called the Air-to-Cloth ratio (A/C expressed in m/s) and is the main design parameter of a PJFF unit. This parameter also represents the filtering velocity of particles through the bags. The lower this velocity is, the higher the needed filtration area is. In other case, a lower filtration area leads to a higher accumulation of dust on bags and to a higher pressure drop [CCSD, 2008].

Air-to-Cloth values are presented in Ref. Box PJFF 1 and are taken from recent industrial examples and manufacturer advices [Termokimik, E-On, Enel, Duke Energy, Hamon, Balcke Dürr]. This ratio ranges between 1.00 10^{-2} and 2.33 10^{-2} m/s.

Once A/C has been chosen, the Net Cloth Area can be calculated through equation 4-1.

$$A_{NC}[m^2] = \frac{\dot{v}_{\lambda,dry,sec}^{flue\ gas}\left[\frac{Nm^3}{s}\right]}{G/C\ \left[\frac{m}{s}\right]}$$

4-1

Because the volumetric flow rate and dust load may fluctuate, A_{NC} should be overestimated. The conversion between Net cloth area to Gross Cloth Area A_{GC} is done by f^{N-G} following equation 4-2:

$$A_{GC}[m^2] = A_{NC}[m^2] \cdot f^{N-G}$$

4-2

f^{N-G} is a function of A_{NC} range and it is summarized in Ref. Box PJFF 2. This value ranges between 1.04 and 2 for smaller Net Cloth Area.

Reference box PJFF2 - Conversion Net to Gross Cloth Area

Level of Net cloth Area (m²)	Multiplicator factor for gross cloth area
< 370	2
< 1115	1,5
< 2230	1,25
< 3350	1,17
< 4460	1,125
< 5580	1,11
< 6690	1,1
< 7810	1,09
< 8920	1,08
< 10040	1,07
< 12270	1,06
< 16730	1,05
> 16730	1,04

Gross cloth area represents the effective filtering media required for a specific dust load. In most combustion power plants, PJFF baghouse structure is divided into several compartments. This number ranges between 2 and 30 depending on the filtration area [E-On, Enel]. This division is done for a better dust loading and flue gas repartition. Maintenance operation and control are easier with subdivision. Compartment area is calculated according to A_{GC} and the number of compartments (equation 4-3).

$$A_{comp}[m^2] = \frac{A_{GC}[m^2]}{nb_{comp}}$$

4-3

Adding extra-compartment to the baghouse structure is also a common practice in large combustion power plant. It has several advantages:

- Cleaning process for PJFF requires from time to time an off-line cleaning in order to remove sticky and sealing dusts.
- One compartment can be off-line while the process is still running.
- Extra-compartments enable to treat a more variable dust charge.
- Maintenance and bag replacement can be done while flue gases are still being filtered.

Extra-compartments increase A_{GC} to a total filtration area A_{tot}according to equation 4-4.

$$A_{tot}[m^2] = A_{GC}[m^2] + A_{comp}[m^2] \cdot nb_{extra-comp} \quad (4\text{-}4)$$

If PJFF is designed without extra-compartments then $A_{tot}[m^2] = A_{GC}[m^2]$

As example, a LCP with a thermal capacity of 1,650 MWth) presents a total average filtration area range from 40,000 to 50,000 square meters.

4.2.2. Fabric filter equipment cost

For the PJFF investment estimations, three equipment costs (C_{equip}^{PJFF} in €$_{2010}$) are distinguished. All calculations are based on the total filtration area A_{tot}. (see fig. 4-4)

4.2.2.1. Baghouse cost $C^{baghouse}$

This cost is the result of compartment cost C^{comp} multiplied by the number of compartment. Compartment division ranges are provided in Ref. box PJFF-3 [Duke Energy, ENEL, EON]. This division could also be calculated from a specific area per compartment.

Reference Box PJFF3 - Filter dimension

Compartment division	1-30
extra compartment	0-2

Each compartment is made of steel and may be improved in stainless steel or insulated to protect the steel from acid gases and rust. Price parameters are available in Ref. box PJFF-4 [US EPA 2002].

Reference box PJFF4 - Price parameters for baghouse compartments - 2010 €

Baghouse type	Component	a (€)	b (€/m2)
Pulse jet (modular)	Basic unit	55 604	124
	SS	26 789	97
	Insulation	3 088	36
Field assembled units	Basic unit	422 647	90
	SS	143 808	34
	Insulation	89 879	10

For pulse jet structure construction, two sets for cost parameters are provided [US EPA, 2002]:

- For little size PJFF unit ($A_{tot} < 4{,}600$ m²), the assembled structure is delivery on site.

- In other cases, the structure is too large and must be field-assembled.

In the second condition, cost parameters are higher.

4.2.2.2. Filtering media cost $\mathbf{C_{tot}^{bag}}$

Type of filtering media defines the bag cost per square meter. The most resistant media is the most expensive one. And naturally, the most impermeable one is also the most expensive. An average cost per square meter is available for PE media from literature $C_{area}^{bag,PE}$. Ref. box PJFF-5 [US EPA, Balcke Durr, Hamon] presents media price factors f^{med} referenced to PE media [US EPA FF lesson 4] and prices per square meter of filtering media are given in Ref. box PJFF-6.

Reference box PJFF5 - Bag cost factors for various materials	
PE	1
CO	1,1
PP	1,2
FG	2,5
NO	5,0
RT	6,3
P8	7,5
TF	9,4

Reference Box PJFF6 - Price Utilities	
PE media price (€/m2)	5-9
Cage price (€/m2 filtering media)	16-25

These costs C_{area}^{bag} are expressed in euro per square meter and are computed with the total filtration area to obtain C_{tot}^{bag} in euro (equations 4-5 and 4-6).

$$C_{area}^{bag}\left[\frac{€}{m^2}\right] = C_{area}^{bag,PE}\left[\frac{€}{m^2}\right] \cdot f^{med} \quad \text{4-5}$$

$$C_{tot}^{bag}[€] = C_{area}^{bag}\left[\frac{€}{m^2}\right] \cdot A_{tot}[m^2] \quad \text{4-6}$$

For coal combustion plant, typical filtering media used is PPS or Ryton. It could be improved in P84 or with a Teflon membrane for high temperature and corrosive fumes.

4.2.2.3. Cage cost $\mathbf{C_{tot}^{cage}}$

Pulse jet cleaning needs counter-current compressed air to remove the dust cake from filter media. In order not to deflate bags, cages are installed inside the bags. Prices per square meter of filtering media are given in Ref. box PJFF-6 (see above).

4.2.3. Fabric filter investment determination

PJFF equipment cost is the sum of the three detailed cost ahead multiplied with a factor including instrumentation (equation 4-7).

$$C_{equip}^{PJFF}[€] = f^{inst} \cdot \left(C^{baghouse}[€] + C_{tot}^{bag}[€] + C_{tot}^{cage}[€]\right) \tag{4-7}$$

This factor is provided by Ref. box PJFF-9 (see above).

Reference Box PJFF9 - Purchase equipment factors

Instrumentation	0.10
Freight	0.05

Total investment for PJFF unit includes the price of PJFF equipments plus direct installation costs and indirect installation costs.

Direct installation cost groups the following items:

- Foundation and support
- Handling and erection
- Electrical
- Piping
- Insulation for ductwork
- Painting

A global direct installation factor f_{inst}^{direct} is given in EPA cost control manual for all of these parameters and is equal to 0,74 (equation 4-8).

$$C_{inst}^{direct}[€] = f_{inst}^{direct} \cdot C_{equip}^{PJFF}[€] \tag{4-8}$$

Indirect installation costs include:

- Engineering
- Construction and filed expense
- Contractor fees
- Start-up

- Performance test
- Contingencies

All of these parameters result in a global indirect installation factor equal to 0.45 of C_{equip}^{PJFF} [US EPA 2002] (equation 4-9).

$$C_{inst}^{indirect}[€] = f_{inst}^{indirect} \cdot C_{equip}^{PJFF}[€] \tag{4-9}$$

Finally, total investment cost is the sum of equations 4-7, 4-8 and 4-9 (equation 4-10)

$$C_{inv}^{PJFF}[€] = C_{equip}^{PJFF}[€] + C_{inst}^{direct}[€] + C_{inst}^{indirect}[€] \tag{4-10}$$

The investment cost calculated by this methodology represents the cost for a new PJFF unit. In case of an existing plant, a retrofit factor should be applied. The value of retrofit factor ranges between 0.3 and 0.5 [Nalbandian, 2006] and is set to 0.4 in this report.

4.2.4. Operating costs

Fixed operating costs, $C_{op,fix}^{PJFF}[€]$ are defined as in chapter 1.2.2. The percentage is fixed to 2%.

Variable operating costs for PJFF units are as follows:

4.2.4.1. Electricity consumption

- Fan utility electricity cost $C_{ut.el}^{fan}$(€/year) (equation 1-6 for the demand in electricity and the electricity cost C_{el}(€/MWh))
- Air compressor utility electricity demand computed by equation 4-11 and cost associated cost $C_{ut.elec}^{air\ comp}$(€/year) computed by equation 4-11-bis.

$$cons_{ut.elec}^{air\ comp}\left[\frac{MWh}{hour}\right] = 0.000375 \cdot \dot{v}_{\lambda,dry,s}^{flue\ gas}\left[\frac{Nm^3}{s}\right] \cdot {}^{c}/_{a} . 3600 \tag{4-11}$$

Where c/a is the ratio between compressed air flow rate to actual air flow rate.

$$C_{ut.elec}^{air\ comp}\left[\frac{€}{year}\right] = cons_{ut.elec}^{air\ comp}\left[\frac{MWh}{hour}\right] . 8{,}760 . CAP[\%] . \frac{1}{100} C_{ut.elec}\left[\frac{€}{MWh}\right] \tag{4-11-bis}$$

4.2.4.2. Bag replacement C_{rep}^{bag}

Bag replacement cost C_{rep}^{bag}(€/year) is as follows.

$$C_{rep}^{bag}\left[\frac{€}{year}\right] = \frac{(1+p[\%])^{n^{bag}}}{(1+p[\%])^{n^{bag}-1}} \cdot p[\%] \cdot \left(C^{bag}[€] + 2{,}15 \cdot A_{tot}[m^2]\right) \tag{4-12}$$

n^{bag} refers to the bag operating lifetime. It depends on the characteristics of flue gas, cleaning frequency and pressure drop. Bag lifetime ranges from 15,000 to more than 40,000 operating hours which may represent from 1 to 5 years depending on capacity factor.

4.2.4.3. By-product disposal or recovery $C_{by-product}$

Ash load is calculated from equation 2-1-14.

The total of fly ash or TSP, or by-product recovered is provided by equation 4-13 (It is important to note, that the fly ash concentrations in eq. 4.13 need to be at actual oxygen concentration, and not at corrected oxygen concentration. This is due to the fact, that the annual flue gas volumes relate to actual oxygen concentration as well and would need to be adjusted to the corrected oxygen concentration otherwise).

$$\begin{aligned} & m_{fly\ ash,abated,PJFF}\left[\frac{t_{ash}}{year}\right] \\ & \quad = \left(load_{fly\ ash,dry,refO2}\left[\frac{mg}{Nm^3}\right] - c_{fly\ ash,load\ emit,refO2}\left[\frac{mg}{Nm^3}\right]\right) \cdot \frac{1}{f_{O2,corr}} \\ & \quad \cdot \dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{year}\right] \cdot 10^{-9} \end{aligned} \tag{4-13}$$

Annual costs are defined according to equation 4-14.

$$C_{fly\ ash\ or\ bp}^{PJFF}\left[\frac{€}{year}\right] = \left(m_{fly\ ash,abated,PJFF}\left[\frac{t_{fly\ ash}}{year}\right] \cdot C_{bp,spe}\left[\frac{€}{t_{fly\ ash}}\right]\right) \tag{4-14}$$

In case of fly ash sold, $C_{bp,spe}$ is negative, indicating a benefit. In case of disposal, $C_{fly\ ash\ or\ bp,spe}$ is positive. According to questionnaires, sale prices range from 0.35 €/t to 6 €/t. In case of disposal, price is around 3.75 €/t. Fly ashes can be reused for cement production and public work applications. The chemical characteristics must satisfy different standards according to applications. It is only in case of unsatisfactory combustion conditions that fly ashes have to be disposed and according to coals burnt.

4.2.5. Efficiency of PJFF

Efficiency of Pulse Jet Fabric Filters is assumed to be higher than 99.0% and may reach 99.99% [US EPA, Hamon, CCSD]. This efficiency is not related to PJFF design but more on operation practice.

The cleaning efficiency of FF results of the low porosity of filtering media. This permeability decreases with the dust accumulation on bags which creates a double filtering layer. This double layer increases the pressure drop across the bags and more fan power is required to counter this air resistance. To control the efficiency and the pressure drop, a compromise should be found regarding the cleaning frequency. Indeed, on the one hand if the cleaning frequency is low, the efficiency will be improved to the detriment of pressure drop and electricity consumption. On the other hand, a too high cleaning frequency reduces the efficiency. The bag lifetime is indeed impacted by cleaning frequency: an increase of compressed air injection results in an increase of maintenance and replacement operations.

To illustrate the influence of main input parameters on output parameters i.e. investment costs and operating costs, the following table is provided.

		Results on output parameters				
		ΔP	$C_{dust}^{bag}\left[\frac{g}{m^2}\right]$	Eff	C_{inv}^{PJFF} [€]	C_{op}^{PJFF} [€])
Increasing input parameters	A/C	↗	↗	↗	↘	↗
	A_{tot}	↘	↘	↘**	↗	↘
	$F_{cleaning}$*	↘	↘	↘	-	↗

Table 4-1 Resulting influence of design parameters

*$F_{cleaning}$ represents the cleaning frequency of filtering media. This input parameter is not developed in the methodology because of the lack of data.

**assuming a medium porosity of filtering media

Table 4-1 presents the result of an increase of the 3 main design parameters on 5 output parameters. One thing that has to be kept in mind that A/C and A_{tot} evolves in opposite way.

4.3. Electrostatic Precipitator

4.3.1. Investment costs

Effective Collecting Plate Area A_{ECP} (m^2) is the main design parameter to scale up an ESP unit. Chapter 4.3.1.1 develops the approaching method to calculate the A_{ECP} and chapter 4.3.1.2 presents the leading equations to estimate costs and investments.

4.3.1.1. Bottom up approach for $\mathbf{A_{ECP}}$ value

The ESP main design parameter is the effective collecting plate area A_{ECP} (m^2). The aim of the method used in US EPA methodology is to approach the Specific Collecting Area SCA (s/m) which is the ratio between A_{ECP}. and $\dot{v}_{\lambda,dry,sec}^{flue\ gas}$. This method requires several input parameters:

- Efficiency $\eta[\%]$
- Temperature $T[K]$
- Mass mean diameter $MMD_{in}[\mu m]$
- The presence or not of Back Corona (BC)
- Several factors (see Ref.box ESP 1 in excel file and table below)

Reference box ESP-1 Values for A ECP determination

Parameter	Value	Unit	Source
Temperature (T)	380-500	[K]	[CCSD 2008 ; Zevenhoven & Kilpinen]
Mass mean Diameter (MMD_{in})	3-21	[µm]	[US EPA, 2002 ; Juda-Rezler, Kowalczyk, 2012]
Sneakage factor (f^{SN})	0,07		[US EPA, 2002]
Raping reentrainment factor (f^{RR})	0,14		[US EPA, 2002]
Most penetrating size (MMD_p)	2	[µm]	[US EPA, 2002]
Rapping puff size (MMD_r)	5	[µm]	[US EPA, 2002]
Free space permittivity (ε_0)	8,845E-12	[F/m]	
Loss factor (f^L)	0,2002		[US EPA, 2002]

Fly ash resistivity depends on numerous parameters such as the chemical composition, the temperature, moisture content...

For simplification reasons, ESP design calculations have been developed for ash resistivity in a range adapted to its correct efficiency, between 10^8 and 2.10^{11} ohm.cm.

Temperature is required to calculate the gas viscosity ν_G and the electric field at sparking E_{bd} (equations 4-15 and 4-16).

$$E_{bd}\left[\frac{V}{m}\right] = 6{,}3.10^5 \cdot \left(\frac{273}{T[K]}\right)^{1{,}65} \tag{4-15}$$

$$\nu_G\left[\frac{kg}{m.s}\right] = 1{,}72.10^{-5} \cdot \left(\frac{273}{T[K]}\right)^{0{,}71} \tag{4-16}$$

Penetration parameter (pe in %) represents the percentage of non-collected dust.

$$pe[\%] = (100 - \eta[\%]) \cdot \frac{1}{100} \tag{4-17}$$

Average electric field E_{avg} is derived from equation 4-18, taking into account the presence or not of BC effect represented by factor f^{BC} .This factor equals 0.57 if no back corona occurs and 0.4 in other case.

$E_{avg}\left[\frac{V}{m}\right] = f^{BC} \cdot E_{bd}\left[\frac{V}{m}\right]$	4-18

From equation 4-17 and loss factor f^{L} value from table 4-1, the number of ESP fields is deduced through the following equation.

$n = ent\left(\frac{ln(pe[\%])}{ln(f^{L})}\right) + 1$	4-19

Then three new parameters are calculated: the average section penetration pe_s (4-20), the section collection penetration pe_c (4-21) and a particle size change factor MMD_{rp} (4-22).

$pe_s = pe^{1/n}$	4-20
$pe_c = \frac{pe_s - f^{L}}{1 - f^{L}}$	4-21
$MMD_{rp}[\mu m] = f^{RR} \cdot (1 - f^{SN}) \cdot (1 - pe_c) \cdot \frac{MMD_r[\mu m]}{pe_s}$	4-22

Each ESP field k is defined from a Mass Mean Diameter $MMD_{k,in}$, which decreases field after field, and a specific collecting area SCA_k, which increases field after field. Equations 4-23 and 4-24 allow respectively the calculation of these parameters.

$For\ k\ from\ 1\ to\ n-1, and\ MMD_{1} = MMD_{in}$ $MMD_{k+1,in}[\mu m] = \frac{\left[MMD_{k,in}[\mu m] \cdot f^{SN} + pe_c \cdot \left((1-pe_c) \cdot MMD_p[\mu m] + pe_c \cdot MMD_{k,in}[\mu m]\right)\right]}{pe_s} + MMD_{rp}[\mu m]$	4-23
$For\ k\ from\ 1\ to\ n$ $SCA_k\left[\frac{s}{m}\right] = -\frac{\nu_G\left[\frac{kg}{m.s}\right]}{\varepsilon_0\left[\frac{C}{V.m}\right]} \cdot (1 - f^{SN}) \cdot \frac{\ln(pe_c)}{\left(E_{avg}\left[\frac{V}{m}\right]\right)^2 \cdot MMD_{k,in}[\mu m] \cdot 10^{-6}}$	4-24

Total SCA is given by summing SCA_k for each ESP field and by multiplying this value with the volumetric gas flow rate, the Effective collecting plate area is obtained.

$$SCA\left[\frac{s}{m}\right] = \sum_{k=1}^{n} SCA_k\left[\frac{s}{m}\right] \tag{4-25}$$

$$A_{ECP}[m^2] = SCA\left[\frac{s}{m}\right]\cdot \dot{v}^{flue\ gas}_{\lambda,dry,sec}\left[\frac{Nm^3}{s}\right] \tag{4-26}$$

4.3.1.2. ESP investment determination

Figure 4-3 summarises the logical flow diagram for the ESP equipment cost determination.

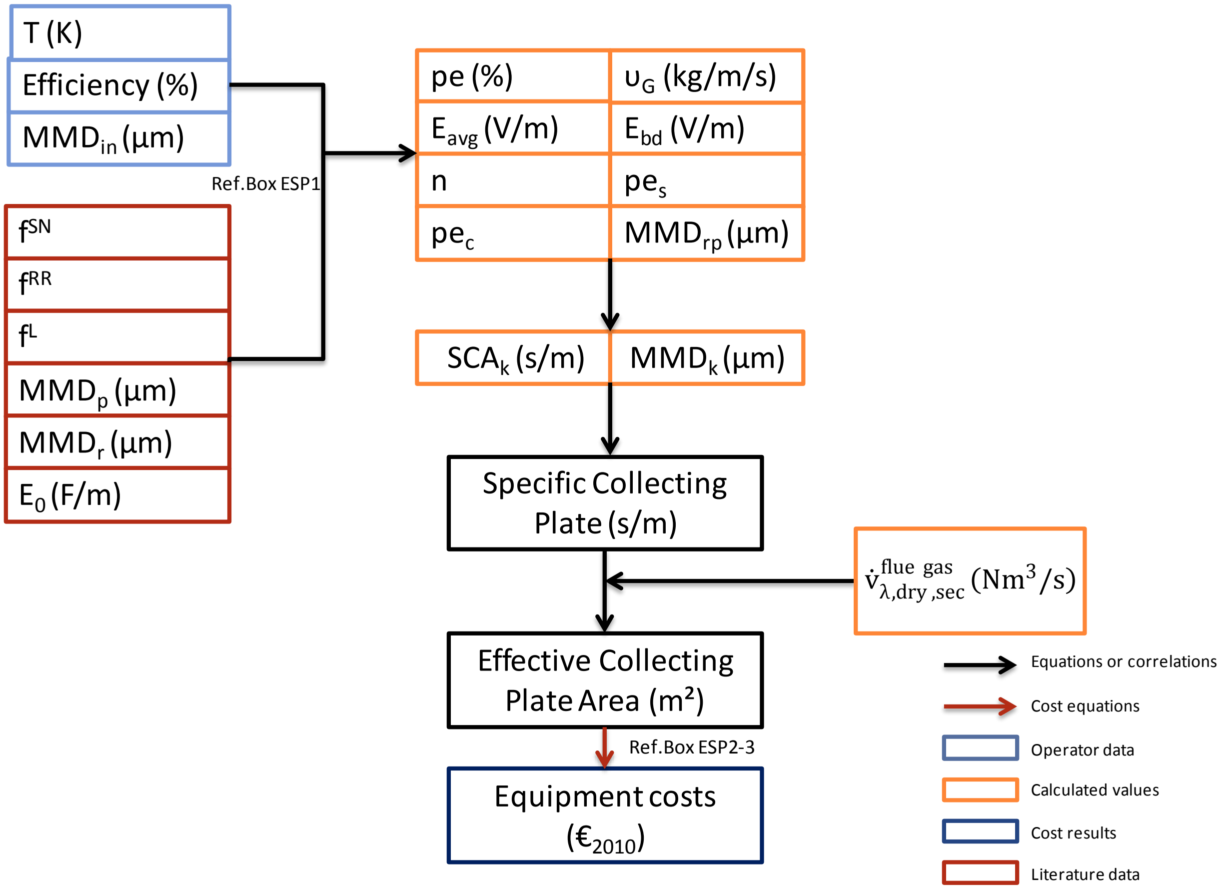


Figure 4-3- Logical flow diagram for ESP equipment cost determination

Main costs for ESP equipment consist in these following parameters:

- ESP casing
- Pyramidal hoppers
- Rigid electrodes and internal collection plates
- Transformer-rectifier (T-R) sets and microprocessor controls

- Rappers
- Stub supports

US EPA develops equations for both costs, depending on effective collecting plate area.

$$C^{ESP,equip}[€] = a \cdot (A_{ECP}[m^2])^b \qquad (4\text{-}27)$$

In equation 4-27, a and b are price parameter given by Ref. box ESP-2. These parameters are also available for “all-standard-option” installation [US EPA, lesson 4].

	Reference box ESP-2 Equipment cost in 2010 €		
	Plate area inferior limit (m2)	a	b
Basic unit	AECP ≤ 4645 m2	3 496	0,6275
	AECP > 4645 m2	549	0,8431
All standard option	AECP ≤ 4645 m2	5 069	0,6276
	AECP > 4645 m2	796	0,8431

According to US EPA, “all-standard-option” includes:

- Inlet and outlet nozzles and diffuser plates
- Hopper auxiliaries/heaters, level detectors
- Weather enclosure and stair access
- Structural supports
- Insulation

If the flue gas is corrosive, compartments have to be protected and be in stainless steel or more resistant material. Material factors $f^{ESP-mat}$ are available in ref.box ESP-3 to derive the calculated value C_{inv}^{ESP}.

Reference box ESP-3 cost using various materials

Material	Factor
Carbon Steel	1
Stainless steel 304	1,3
Stainless steel 316	1,7
Carpenter 20 CB-3	1,9
Monel-400	2,3
Nickel-200	3,2
Titanium	4,5

A global installation factor $f^{ESPinst}$ is given in EPA cost control manual in ref. Box ESP-5. $C^{ESP,inst}$ is given by equation 4-28.

$$C_{equip}^{ESPcomplete}[€] = f^{inst} \cdot (C^{ESP\ equip}[€] \cdot f^{ESP-mat} + C^{SO3}[€]) \qquad (4\text{-}28)$$

Total investment for ESP unit includes the price of ESP equipment plus the following installation parameters:

- Foundation and support
- Handling and erection
- Electrical
- Piping
- Insulation for ductwork
- Painting

A global direct installation factor f_{inst}^{direct} is given in EPA cost control manual for all of these parameters and is equal to 0,67 (equation 4-28bis).

$$C_{inst}^{direct}[€] = f_{inst}^{direct} \cdot C_{equip}^{ESP,complete}[€] \tag{4-28 bis}$$

Indirect installation costs include:

- Engineering
- Construction and filed expense
- Contractor fees
- Start-up
- Performance test
- Contingencies

All of these parameters result in a global indirect installation factor equal to 0.57 of $C_{equip}^{ESP,complete}$ [US EPA 2002] (equation 4-28 ter).

$$C_{inst}^{indirect}[€] = f_{inst}^{indirect} \cdot C_{equip}^{ESP,complete}[€] \tag{4-28 ter}$$

Finally, total investment cost is the sum of equation 4-28, 4-28 bis and 4-28 ter (equation 4-28 four).

$$C_{inv}^{ESP}[€] = C_{equip}^{ESP\ complete}[€].r + C_{inst}^{direct}[€] + C_{inst}^{indirect}[€] \tag{4-28 four}$$

In case of an existing plant, a retrofit factor should be applied. The value of retrofit factor ranges between 0.3 and 0.5 [Nalbandian, 2006] and is set to 0.4 in this report.

4.3.2. Operating costs

Fixed operating costs defined in chapter 1.1.2.

Operating Costs are as follows:

4.3.2.1. Electricity consumption

- Fan utility electricity consumption (MWh/h) (equation 1-6)
- ESP power requirement P^{ESP}(MWh/h) (equation 4-29).

$$P^{ESP}\left[\frac{MWh}{hour}\right] = 2{,}09.10^{-5} \cdot A_{ECP}[m^2] \tag{4-29}$$

4.3.2.2. SO$_3$ consumption C_{SO_3}

SO$_3$ is injected in ESP in order to decrease the ash resistivity. Ash resistivity depends on the temperature, their moisture content, their sulphur content and their chemical composition. Ash particles can reach high resistivity values [US EPA, lesson 3]. In that case, the electric power used to collect particles become insufficient and the ESP has to be enlarged. This phenomenon is called the back corona. In order to avoid the back corona effect, SO$_3$ is injected in the fumes before the ESP [Evonik, 2011].

Due to too complex ash resistivity calculation, simple hypotheses have been developed. These hypotheses have been discussed by industrial experts [EDF 2013, HAMON2013].

In the EGTEI methodology, we assume that only coal ashes could lead to the back corona effect. Indeed, we also assume that back corona effect in case of biomass co-firing is avoided by the moisture content of biomass.

For precautionary reasons, in the EGTEI methodology, SO$_3$ may be systematically injected to prevent coal ash high resistivity.

The SO$_3$ consumption cost is given by the following equation.

$$C_{SO_3}\left[\frac{€}{year}\right] = \frac{Q_{SO_3}\left[\frac{kg}{hour}\right]}{1000} \cdot \frac{CAP}{100} . 8{,}760\left[\frac{hours}{year}\right] \cdot C_{SO_3,m}\left[\frac{€}{ton}\right] \tag{4-30}$$

Ref. box ESP 6 gives a range of SO$_3$ consumption.

Reference Box ESP-6 SO3 conditionning	
SO3 injection rate (kg/h)	10-80
Sulfur cost (€/t)	70

4.3.2.3. By-product disposal or recovery $C_{by-product}$

Ash load is calculated from equation 2-1-14.

The total of fly ash or TSP, or by-product recovered is provided by equation 4-31 (It is important to note, that the fly ash concentrations in eq. 4-31 need to be at actual oxygen concentration, and not at corrected oxygen concentration. This is due to the fact, that the annual flue gas volumes relate to actual oxygen concentration as well and would need to be adjusted to the corrected oxygen concentration otherwise).

$$\begin{aligned} m_{fly\ ash,abated,ESP}\left[\frac{t_{ash}}{year}\right] &= \left(load_{fly\ ash,dry,refO2}\left[\frac{mg}{Nm^3}\right] - c_{fly\ ash,load\ emit,dry,ref02}\left[\frac{mg}{Nm^3}\right]\right)\cdot\frac{1}{f_{O2,corr}} \\ &\cdot \dot{v}^{flue\ gas}_{\lambda,dry,year}\left[\frac{Nm^3}{year}\right]\cdot 10^{-9} \end{aligned} \tag{4-31}$$

Annual costs are defined according to equation 4-32.

$$C^{ESP}_{fly\ ash\ or\ bp}\left[\frac{€}{year}\right] = \left(m_{fly\ ash,abated,ESP}\left[\frac{t_{ash}}{year}\right]\cdot C_{bp,spe}\left[\frac{€}{t_{bp}}\right]\right) \tag{4-32}$$

In case of fly ash sold, $C_{bp,spe}$ is negative, indicating a benefit. In case of disposal, $C_{bp,spe}$ is positive. According to questionnaires, sale prices range from 0.35 €/t to 6 €/t. In case of disposal, price is around 3.75 €/t. Fly ashes can be reused for cement production and public work applications. The chemical characteristics must satisfy different standards according to applications. It is only in case of unsatisfactory combustion conditions that fly ashes have to be disposed and according to coals burnt.

4.4. Biomass co-firing and liquid fuels

For biomass co-firing, moisture content of biomass is an interesting characteristic to enhance the ESP efficiency. Compared to coal, biomass has high moisture content (20-40%) which is useful for preventing back corona effect in ESP. Even a small percentage of co-fired biomass increases the dust removal efficiency [Jedrusik, Swierczok, 2012].

This characteristic avoids the use of SO_3 injection and serves as a guarantee for an operating resistivity. Moreover, biomass co-firing may develop sticky and agglomerated dust which is beneficial for dust collecting velocity in ESP.

On contrary, these characteristics of biomass are harmful for fabric filtering media. If PJFF is used, biomass particles may seal irreparably the filtering media in a non-reversible way. Then, bag replacement occurs more frequently and this increases maintenance cost.

For the same reason, heavy liquid fuel requires an ESP to eliminate combustion particles. Unburned combustion particles may plug the filtering media as well as biomass particles [Hamon, 2013].

5. Evaluation of SO_2 abatement techniques

Three abatement techniques are considered:

- Wet flue gas desulphurisation (FGD) with limestone forced oxidation (LSFO FGD). This technique is frequently used in coal combustion plants. According to [Nalbandian 2006], it represents 89% of the market share for these coal combustion plants. There is rare information for plants using liquid fuels. The LFSO FGD enables to produce commercial grade gypsum which can be used for plaster production, cement production as well as other applications if technical specifications are achieved (however the market is saturated and the opportunities of valorisation decrease according to information provided). LSFO FGD can be used theoretically for plants of any size but is generally used for the largest ones, due to economic reasons [Nalbandian 2006]. LSFO FGD can be installed for coals of all sulphur level, including low sulphur coals (less than 1% S) but are typically used for coals with medium and high sulphur content (> 2% S) [Sargent 2007]. Efficiency ranges between 85 and 98% [EU Commission 2006a]. 99% is also provided [Sargent 2007], however sulphur level lower than 0.017 kg SO_2/GJ is a practical outlet limitation with the lowest sulphur content coals (about 50 mg SO_2/Nm^3, considering a coal average LHV (Lower Heating Value) of 29 GJ/t coal).
- Lime spray dryer (LSD) FGD. This technique most often employs lime in a semi-dry form. According to [Nalbandian 2006], it represents 8% of the market share for plants using coal. No information is available for plants using liquid fuels. Waste gas stream is brought into contact with lime in a spray dryer. A fabric filter is used just after, to remove dust and calcium sulphates and sulphites formed during desulphurisation. Lime spray dryers can be used for plants theoretically of any size. However, they are mainly observed for plants lower than 250 MWe capacity (and at maximum 350 MWe) [Nalbandian 2006]. By-products have fewer uses than those from LSFO FGD, due to their properties (mixture of fly ashes and by-products).

  LSD FGD is used for low and medium sulphur coal (< 1.5% S) [Nalbandian 2006]. Efficiency ranges between 70 to 90 % [EU Commission, 2006a]. 95% is also provided by Sargent 2007], depending on the ratio Ca/S.
- Dry FGD. They are several arrangements for dry FGD [EU Commission, 2006a]. The duct sorbent injection FGD is considered for cost estimation (DSI FGD). Several types of reagent can be used such as dry hydrated lime ($Ca(OH)_2$), quick lime (CaO) and sodium Bicarbonate ($NaHCO_3$). Due to high amount of un-reacted sorbent, recycling of products is often operated several times. The specific surface of reagent is crucial for the efficiency of the process. Reagent prices depend on this characteristic. According to [EU commission, 2006a], the existing ESP can be kept in operation in case of retrofit (It is explained, that the way an ESP is able to collect 100 times more product and fly ash than at a single throughput to a duct injection process, is not well understood) and sorbents can be injected in the ducts before.

  Dry FGD efficiency ranges from 50 to 80 % according to [EU Commission, 2006a]. Manufacturers of these technologies announce larger efficiencies up to 99 %, depending on sorbent quality (specific surface), arrangement made and excess of sorbent used.

In the cost methodology developed by TFTEI, the use of a fabric filter is considered. It is installed after the ESP. In this document, cost estimation is developed for the use of lime and sodium bicarbonate.

5.1. LSFO FGD

5.1.1. Investment costs

Annex D presents capital costs available in the literature and collected from the questionnaires received from plant operators in the EU. The cost function developed by US EPA from a study carried by Sargent and Lundy (Sargent 2010b), has been selected and adapted to represent the evolution of investment C_{inv}^{LSFO} according to the size of the combustion plant and its characteristics.

The US EPA cost function has been adapted to accept units of this report (international system of units) and express costs in Euro.

The investment C_{inv}^{LSFO} includes three groups of cost:

1 - The unit cost C_{equip}^{LSFO} itself:

C_{equip}^{LSFO} includes all equipment, installation, buildings, foundations, electrical equipment, minor waste water treatment plant and takes into account the retrofit difficulty. The cost function to derive C_{equip}^{LSFO} includes different modules:

- absorber unit cost C_{abs}^{LSFO},
- reagent preparation unit cost C_{reag}^{LSFO},
- waste handling unit cost C_{wh}^{LSFO},
- base balance of plant costs including: ID or booster fans, new wet chimney, piping, ductwork, minor waste water treatment, etc. C_{bbp}^{LSFO}.

C_{equip}^{LSFO} which is the sum of the four previous components, is the total unit cost.

The US EPA model assumes a wet chimney and no gas/gas heater after the SO_2 absorber. According to [Nalbandian 2006] and confirmed by the TFTEI/EGTEI LCP working group, this is presently a common practice. After the wet absorber, the chimney exit temperature can be as low as 50 °C. The use of a wet chimney instead of a dry chimney avoids installation of a gas/gas heater. The main advantages of omitting the gas/gas heater are provided by [Nalbandian 2006]:

- Cost,
- Removal of a maintenance intensive piece of equipment,
- Lower additional auxiliary power consumption.

A wet chimney is however expensive as it must be corrosion resistant (and has to be built in alloy, alloy carbon steel on contrary to a dry chimney constructed in traditional carbon steel [Sargent 2007]). With a wet chimney, the risk of a visible plume is higher than with a dry chimney.

2 – Additional indirect cost C_{inst}^{ind} for:

- Engineering and construction management,
- Labour cost for construction (labour adjustments for 6 x 10 hour shift premium, per diem),
- Contractor profit and fees.

Those items are assumed to represent up to 30% of C_{equip}^{LSFO}, according to [Sargent 2010b]. 30% has been taken into account as default value in the EXCEL file.

3 - Owner's home office C_{ho}^{LSFO} costs:

Owner's home office costs (owner's engineering, management and procurement) assumed to be 5% of the sum C_{equip}^{LSFO} and C_{inst}^{ind} by [Sargent 2010b]. (This subtotal C_{equip}^{LSFO} and C_{inst}^{ind} is the capital, engineering and construction cost).

The US EPA model also includes allowance for funds used during construction (AFUDC) and owner costs. The AFUDC is based on a three year engineering and construction cycle. This component of costs is not taken into account in the EGTEI methodology.

Investment C_{inv}^{LSFO} is calculated by the following formula.

$$C_{inv}^{LSFO}[€\,] = C_{equip}^{LSFO}[€\,] + C_{inst}^{ind}[€\,] + C_{ho}^{ind}[€\,] \tag{5-1}$$

The cost functions for the different modules of the LSFO FGD, are as follows:

- C_{abs}^{LSFO} : absorber unit cost.

$$C_{abs}^{LSFO}[€\,] = 4{,}672{,}327 \cdot r \cdot \left(cf \cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.6} . \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right]\right)^{0.02} \cdot (bs\ [MWe])^{0.716} \tag{5-2}$$

cf: coal factor: 1 for bituminous, 1.07 for lignite

The load of SO$_2$ is calculated by formula 5-2 bis.

$$load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right] = \frac{\dfrac{M_{SO2}\left[\frac{kg}{kmol}\right]}{Ms\left[\frac{kg}{kmol}\right]} \cdot x_S[\%] \cdot 10\ \left[\frac{kg\ S}{kg\ fuel}\right] \cdot \left[\frac{100 - x_{ash\,[\%]} - x_{moist}[\%]}{10000}\right] \cdot \left[\frac{100 - x_{S\,retainedinbo}[\%]}{100}\right]}{LHV^{fuel}\ \left[\frac{MJ}{kg\ fuel}\right]} \cdot 1000\left[\frac{MJ}{GJ}\right] \tag{5-2 bis}$$

- C_{reag}^{LSFO}: reagent preparation unit cost.

$$C_{reag}^{LSFO}[€\ 2010] = 528{,}986 \cdot r \cdot \left(\left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right] \cdot GHR\left[\frac{GJ}{kWh}\right] \right)^{0.3} \right) \cdot (bs[MWe])^{0.716} \tag*{5-3}$$

- C_{wh}^{LSFO}: waste handling unit cost.

$$C_{wh}^{LSFO}[€\ ] = 625{,}373 \cdot r \cdot \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right] \cdot GHR\left[\frac{GJ}{kWh}\right] \right)^{0.45} \cdot (bs[MWe])^{0.716} \tag*{5-4}$$

- C_{bbp}^{LSFO}: base balance of plant costs.

$$C_{bbp}^{LSFO}[€\ ] = 3{,}442{,}158 \cdot r \cdot \left(cf \cdot GHR\left[\frac{GJ}{kWh}\right] \right)^{0.4} \cdot (bs[MWe])^{0.716} \tag*{5-5}$$

cf : coal factor: 1 for bituminous, 1.07 for lignite

The total equipment cost including retrofit factor $C_{equip}^{LSFO}[€\ ]$ is the following.

$$C_{equip}^{LSFO}[€\ ] = C_{abs}^{LSFO}[€\ ] + C_{reag}^{LSFO}[€\ ] + C_{wh}^{LSFO}[€\ ] + C_{bbp}^{LSFO}[€] \tag*{5-6}$$

$C_{inst}^{ind}[€]$ includes engineering and construction management cost, labour cost for construction and contractor profit and fees and is estimated as a percentage of C_{equip}^{LSFO}.

$$C_{instal}^{ind}[€\ ] = \frac{f_{ec}\,[\%]}{100} \cdot C_{equip}^{LSFO}[€] \tag*{5-7}$$

f_{ec}: factor for engineering and construction management cost, for labour and contractor profit. This factor is taken by default at 30 % (30 % as assumed by [Sargent 2010b]).

$C_{ho}^{ind}[€]$, owner's home office costs are estimated as follows.

$$C_{ho}^{ind}[€] = \frac{f_{ho}[\%]}{100} \cdot (C_{equip}^{LSFO}[€\ ] + C_{instal}^{ind}[€\ ]) \tag*{5-8}$$

f_{ho}: factor for owner home office (5% of the capital, engineering and construction cost $[C_{equip}^{LSFO} + C_{inst}^{ind}]$)

5.1.2. Operating costs

Annual operating costs (C_{op}) include fixed operating and maintenance costs ($C_{op,fix}$) and variable operating costs ($C_{op,var}$).

Fixed operating costs

As presented in chapter 1.1.2, fixed operating costs are calculated as a percentage of the unit cost $C_{equip}^{LSFO}[€]$. Based on [US EPA 2010], the factor OM_{fix} in % of the investment is a function of the size of the plant. From data presented [US EPA 2010], the following function can be derived.

$OM_{fix}[\%] = 0.1324 \cdot bs^{-0.284}\ [MWth] \cdot 100$	5-9

Variable operating costs

For this technique, variable operating costs include auxiliary power, limestone consumption cost, and waste handling cost. Make-up water cost [Sargent 2010b] has been neglected, as quite low compared to other cost items.

5.1.2.1. Auxiliary electricity

The auxiliary power penalty ranges from 1 to 2% of the plant capacity according to [Sargent 2010b] and experts of the EGTEI LCP working group. Higher power penalty can be observed when a gas/gas heater is present [Nalbandian 2006, EGTEI experts].

The penalty has been derived from data of power penalty provided by [US EPA 2010] according to the heat rate (electricity penalty of 1.5% for a gross heat rate of 0.0095 GJ/kWh, 1.67% for a heat rate of 0.0106 GJ/kWh, 1.84% for a heat rate of 0.0116 GJ/kWh). A function can be derived in which η^{gross} is used.

Equation 5-10 provides annual electricity consumption cost for LSFO FGD.

$C_{elec}^{LSFO}[€\ /year] = bs\ [MWth] \cdot \frac{\eta^{gross}\ [\%]}{100} \cdot \frac{CAP\ [\%]}{100} \cdot 8{,}760 \left[\frac{h}{year}\right] \cdot \left(-0.049 \cdot \frac{\eta^{gross}\ [\%]}{100} + 3.3536\right) \cdot \frac{1}{100} \cdot C_{elec,spe} \left[\frac{€}{MWh}\right]$	5-10

CAP: capacity factor in % (equation 2.1-5)

$C_{elec,spe}$: specific cost of electricity in €/MWh

5.1.2.2. Limestone consumption

Limestone demand depends on the efficiency required and the load of SO_2. Hypothesis taken into account to derive the limestone consumption according to the efficiency of the desulphurisation required, is as follows:

At stoichiometry, 1 mole $CaCO_3$ abates 1 mole SO_2. In terms of mass, the molar ratio Ca/S = 1, accounts to $CaCO_3$ consumption of 1.5625 t $CaCO_3$/t SO_2 abated. The current EGTEI methodology rates have been considered correct by the EGTEI LCP subgroup. The rate of use of limestone is as follows:

Table 5.1: limestone consumption factor (current EGTEI methodology)

Efficiency of SO_2 removal η	t $CaCO_3$/t SO_2 abated	Ratio Ca/S abated
85.0%	1.41	0.90
90.0%	1.48	0.95
95.0%	1.59	1.02

According to its characteristics, limestone purity is about 95 to 97%. The purity has been introduced in the limestone demand.

The specific limestone demand γ_{CaCO3}^{S}per t SO_2 abated is derived as follows, according to the LSFO FGD efficiency required and limestone purity (equation 5.11).:

$$\gamma_{CaCO3}^{S}\left[\frac{t\ CaCO3}{t\ SO2\ abated}\right]=\frac{\left(1.8\cdot\frac{\eta^{LSFO}[\%]}{100}-0.1267\right)}{\frac{PUR_{CaCO3}\ [\%]}{100}} \qquad \text{5-11}$$

γ_{CaCO3}^{S}: specific $CaCO_3$ demand in t $CaCO_3$/t SO_2 removed

η^{LSFO}: removal efficiency required for LSFO

PUR_{CaCO3}: purity of the reagent $CaCO_3$ in %

The amount of SO_2 abated is computed by equation 5-12 (It is important to note, that the SO_2 concentrations in eq. 5-12 need to be at actual oxygen concentration, and not at corrected oxygen concentration. This is due to the fact, that the annual flue gas volumes relate to actual oxygen concentration as well and would need to be adjusted to the corrected oxygen concentration otherwise):

$$m_{SO2\ abated}^{LSFO\ FGD}\ \left[\frac{t_{\ SO2}}{year}\right] = \left(c_{SO2,in,ref\,O2}\left[\frac{mg}{Nm^3}\right] - c_{SO2,load\ emit,ref\,O2}\left[\frac{mg}{Nm^3}\right]\right)\cdot\frac{1}{f_{O2,corr}} \cdot \dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{year}\right]\cdot 10^{-9}\left[\frac{t}{mg}\right] \tag{5-12}$$

Equation 5-12 bis can also be used:

$$m_{SO2\ abated}^{LSFO}\left[\frac{t_{SO2}}{year}\right] = load_{SO2}\left[\frac{kg_{\ SO2}}{GJ}\right]\cdot cons_e^{fuel}\left[\frac{GJ}{h}\right]\cdot\ 0.001\left[\frac{tons}{kg}\right]\cdot\frac{CAP\ [\%]}{100} \cdot 8{,}760\,\frac{\eta^{LSFO}}{100} \tag{5-12 bis}$$

Cost of limestone consumption is provided by equation 5-13:

$$C_{CaCO3}^{LSFO}\left[\frac{€}{year}\right] = \gamma_{CaCO3}^{s}\left[\frac{t_{\ CaCO3}}{t_{\ SO2}}\right]\cdot m_{SO2\ abated}^{LSFO\ FGD}\left[\frac{t_{\ SO2}}{year}\right]\cdot\ C_{CaCO3,spe}\left[\frac{€}{t_{\ CaCO3}}\right] \tag{5-13}$$

$CaCO_3$ prices depend on quantity bought and quality. From questionnaires, prices range from: 11 to 16 €/t $CaCO_3$ in a 2465 MWth plant and 32 to 36 €/t $CaCO_3$ in a 630 MWth plant and 40 €/t $CaCO_3$ in another 630 MWth plant for similar $CaCO_3$ purity (94 % to 96 %)

5.1.2.3. By-product, disposal or recovery

In ideal situation, gypsum can be produced and sold. Commercial grade gypsum can be used in wallboard, cement or plaster manufacturing, also soil conditioner. Special specifications have to be observed such as chlorine content, purity, colour... [Sargent 2007]. Often, different sources of coals can be envisaged in plants. Some coals do not enable to be in the specifications.

Presently, sale price of gypsum produced is very low and operators do not expect higher costs (EGTEI LCP expert group). This is confirmed by [Nalbandian 2006] who also explains that sale prices can be negative when transport costs are included.

If not sold, by-products have to be disposed in correct conditions.

According to oxidation efficiency, $CaSO_3.½H_2O$ and/or $CaSO_4,2H_2O$ are produced. $CaSO_4,2H_2O$ is a saleable product. It is assumed a by-product with a humidity of 10 % [EU Commission, 2006a] (This explains the factor 1.91 in equation 5.14).

γ_{bp}^{s}: specific by-product production in ton by-product (bp)/ton SO_2 removed.

$$\gamma_{bp}^{s}\left[\frac{t_{bp}}{t_{SO2\ abated}}\right] = \gamma_{CaCO3}^{s}\left[\frac{t\ CaCO3}{t\ SO2}\right] \cdot \left(\frac{100 - PUR_{CaCO3[\%]}}{100} \cdot 1 + \frac{PUR_{CaCO3[\%]}}{100} \cdot 1.91\right)\left[\frac{t_{bp}}{t_{CaCO3}}\right] \tag{5-14}$$

Cost of waste disposal or gypsum sold is given by equation 5-15:

$$C_{bp}^{LSFO}\left[\frac{€}{year}\right] = m_{SO2\ abated}^{LSFO\ FGD}\left[\frac{t_{SO_2}}{year}\right] \cdot \gamma_{bp}^{s}\left[\frac{t_{bp}}{t_{SO_2 abated}}\right] \cdot C_{bp,spe}\left[\frac{€}{t_{bp}}\right] \tag{5-15}$$

In case gypsum is sold, $C_{bp,spe}$ is negative, indicating a benefit. In case of disposal, $C_{bp,spe}$ is positive. Price of sold gypsum can be low due to saturation of the market. Questionnaires provide a range from 0.15 to 2 €/t by-product sold. When by-products are disposed, disposal prices depend on the waste disposal treatment. Landfill or other treatments such as incineration can be used. By-product disposal prices range from 0.33 to 89 €/t by-product, according to the information obtained.

5.1.3. Adaptation of costs for installations consuming heavy fuel oil

As carried out in the current EGTEI methodology, the investment function derived for coals is adapted for liquid fuels according to the relative flue gas volume, considering 1 for hard coal.

The average flue gas volume for coal at 6% O_2 is 358 Nm^3/MJ.

The average flue gas volume for heavy fuel oil at 3% O_2 is 285 Nm^3/MJ.

An average factor of 0.8 can be derived from those data. The investment functions presented above for coal are adapted with this factor, as follows.

The cost functions for the different modules of the LSFO FGD for heavy fuel oils, are as follows:

- C_{abs}^{LSFO} : absorber unit cost

$$C_{abs}^{LSFO}[€\,] = 3{,}737{,}862 \cdot r \cdot \left(\left(GHR\left[\frac{GJ}{kWh}\right]\right)^{0.6}\right) \cdot \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right]\right)^{0.02} \cdot (bs\;[MWe])^{0.716} \tag{5-16}$$

- C_{reag}^{LSFO}: reagent preparation unit cost

$$C_{reag}^{LSFO}[€\,2010] = 423{,}189 \cdot r \cdot \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right] \cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.3} \cdot (bs[MWe])^{0.716} \tag{5-17}$$

- C_{WH}^{LSFO}: waste handling unit cost

$$C_{WH}^{LSFO}[€\,] = 500{,}298 \cdot r \cdot \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right] \cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.45} \cdot (bs[MWe])^{0.716} \tag{5-18}$$

- C_{bbp}^{LSFO}: base balance of plan costs

$$C_{bbp}^{LSFO}[€\,] = 2{,}753{,}726 \cdot r \cdot \left(GHR\left[\frac{GJ}{kWh}\right]\right)^{0.4} \cdot (bs[MWe])^{0.716} \tag{5-19}$$

The total equipment cost including retrofit factor $C_{equip}^{LSFO}[€\,]$ is provided by equation 5-6.

Other equations for total investment and operating costs do not change.

5.2. Lime spray dryer FGD (LSD FGD)

In lime spray dryer, either hydrated lime ($Ca(OH)_2$) or quicklime (CaO) can be used. Quicklime is more often used for economical reasons [Sargent 2007]. Lime is slaked to produce a lime milk. Lime milk is atomised to a cloud of fine droplets in the spray dry absorber. Water is evaporated by the heat of the flue gas [EU Commission, 2006a]. Wastes are dry because all the water is evaporated. The residue obtained is a mixture of calcium sulphite, calcium sulphate and fly ash. The presence of fly ash depends on the fact a deduster is maintained before the absorber (in case of retrofit indeed, if the plant is equipped with an ESP, it can be maintained). By-products obtained are not attractive commercially [EU Commission, 2006a].

5.2.1. Investment costs

The cost functions derived assume the use of ball mill slaker [EPA 2010]. It is unclear in this reference, if the fabric filter is included or not in the investment function. It is assumed to be included.

Investment C_{inv}^{LSD} is calculated by the following formula:

$$C_{inv}^{LSD}[€] = C_{equip}^{LSD}[€] + C_{inst}^{LSD}[€] + C_{ho}^{LSD}[€] \tag{5-20}$$

The cost functions for the different modules of the LSD FGD, are as follows [Sargent 2010c]:

- C_{abs}^{LSD} : absorber unit cost

$$C_{abs}^{LSD}[€] = 4{,}767{,}876 \cdot r \cdot \left(cf \cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.6} \cdot \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right]\right)^{0.01} \cdot (bs\ [MWe])^{0.716} \tag{5-21}$$

cf: coal factor: 1 for bituminous, 1.07 for lignite

GHR: gross heat rate expressed in GJ/kWh

$load_{SO2}$: load of SO_2 expressed in kg_{SO2}/GJ (calculated by formula 5-2 bis)

- $C_{reag,wh}^{LSD}$: reagent preparation and waste handling unit cost

$$C_{reag,wh}^{LSD}[€] = 486{,}982 \cdot r \cdot \left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right] \cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.2} \cdot (bs[MWe])^{0.716} \tag{5-22}$$

- C_{bbp}^{LSD}: base balance of plan costs

$$C_{bbp}^{LSD}[€] = 2{,}723{,}053 \cdot r \cdot \left(cf \cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.4} \cdot (bs[MWe])^{0.716} \tag{5-23}$$

cf : coal factor: 1 for bituminous, 1.07 for lignite

The total equipment cost including retrofit factor $C_{equip}^{LSD}[€]$ is:

$$C_{equip}^{LSD}[€] = C_{abs}^{LSD}[€] + C_{reag,wh}^{LSD}[€] + C_{bbp}^{LSD}[€] \tag{5-24}$$

$C_{inst}^{ind}[€]$ includes engineering and construction management cost, labour cost for construction and contractor profit and fees and is estimated as a percentage of C_{equip}^{LSFO} .

$C_{instal}^{ind}[€\,] = \frac{f_{ec}\,[\%]}{100} \cdot\, C_{equip}^{LSD}[€]$	5-25

f_{ec}: factor for engineering and construction management cost, for labour and contractor profit. This factor is taken by default at 30 % (30 % as assumed by [Sargent 2010c]).

$C_{ho}^{ind}[€]$, owner's home office costs are estimated as follows:

$C_{ho}^{ind}[€] = \frac{f_{ho}[\%]}{100} \cdot (C_{equip}^{LSD}[€\,] +\, C_{instal}^{ind}[€\,])$	5-26

f_{ho} : factor for owner home office (5% of the capital, engineering and construction cost $\left[C_{equip}^{LSFO} + Cinstind\right)$

5.2.2. Operating costs

Annual operating costs (C_{op}) include fixed operating and maintenance costs ($C_{op,fix}$) and variable operating costs ($C_{op,var}$).

Fixed operating costs

As presented in chapter 1.1.2, fixed operating costs are calculated as a percentage of the unit cost $C_{equip}^{LSD}[€]$. Based on [EPA 2010], the factor OM_{fix} in % of the investment is a function of the size of the plant. From data presented, the following function can be derived:

$OM_{fix}[\%] = 0.1688 \cdot bs^{-0.297}\ [MWth] \cdot 100$	5-27

Variable operating costs

For LSD FGD, variable operating costs include auxiliary power, lime consumption cost, waste disposal cost.

5.2.2.1. Auxiliary electricity

The auxiliary power penalty ranges from 1 to 2% of the plant capacity according to [Sargent 2010c]. Higher power penalty can be observed when a gas/gas heater is used [Nalbandian 2006, EGTEI experts].

The penalty has been derived from data of power penalty provided by [US EPA 2010] according to the heat rate (electricity penalty of 1.2% for a heat rate of 0.0095 GJ/kWh, 1.33% for a heat rate of 0.0106 GJ/kWh, 1.47% for a heat rate of 0.0116 GJ/kWh). A function can be derived in which η^{gross} is used.

Equation 5-28 provides annual electricity consumption cost for LSD FGD.

$$C_{elec}^{LSD}[€\ /year] = bs\ [MWth]\ \cdot\ \frac{\eta^{gross}\ [\%]}{100} \cdot\ \frac{CAP\ [\%]}{100} \cdot 8{,}760\ \left[\frac{hours}{year}\right] \cdot \left(-0.039\ \cdot\ \frac{\eta^{gross}\ [\%]}{100} + 2.6553\right) \cdot \frac{1}{100} \cdot\ C_{elec,spe}\left[\frac{€}{MWh}\right] \tag{5-28}$$

CAP: capacity factor in % (equation 2.1-5)

$C_{elec,spe}$: specific cost of electricity in €/MWh

5.2.2.2. Lime consumption

Lime demand depends on the efficiency required and the load of SO_2. Hypothesis taken into account to derive the lime consumption according to the desulphurisation efficiency required is as follows: at stoichiometry, 1 mole CaO abates 1 mole SO_2. In terms of mass, the molar ratio Ca/S = 1, accounts to CaO consumption of 0.875 t CaO/t SO_2 abated. [ADEME 2000] provides the following data:

Table 5.2: lime consumption factor for LSD FGD

Efficiency of SO_2 removal η	t CaO/t SO_2 abated	Ratio Ca/S abated
80.0%	0.875	1
90.0%	1.312	1.5
92.0%	1.750	2

According to its characteristics, lime purity is about 95 to 97%. The purity has been introduced in the lime demand. A function has been derived to express the consumption according to the LSD FGD efficiency (equation 5.29).

The specific lime demand γ_{CaO}^{s} per t SO_2 abated is derived as follows, according to the LSD FGD efficiency required and lime purity:

$$\gamma^{s}_{CaO}\left[\frac{t\ CaO}{t\ SO2\ abated}\right] = \frac{\left(6.3508\ \cdot\ \frac{\eta^{LSD}[\%]}{100} -\ 4.2339\right)}{\frac{PUR_{CaO}\ [\%]}{100}} \tag{5-29}$$

γ^{s}_{CaCO3}: specific CaO demand in t CaO/t SO_2 removed

η^{LSD}: removal efficiency required for LSD FGD

PUR_{CaO}: purity of the reagent CaO in %

(For efficiency lower than 70%, the specific lime demand is fixed to the lime demand obtained for an efficiency of 70%, due to possible negative figures in the equation 5-29)

The amount of SO_2 abated is computed by equation 5-30 (It is important to note, that the SO_2 concentrations in eq. 4-30 need to be at actual oxygen concentration, and not at corrected oxygen concentration. This is due to the fact, that the annual flue gas volumes relate to actual oxygen concentration as well and would need to be adjusted to the corrected oxygen concentration otherwise):

$$\begin{aligned} m^{LSD\ FGD}_{SO2\ abated}\ \left[\frac{t_{\ SO2}}{year}\right] &= \left(c_{SO2,in,refO2}\left[\frac{mg}{Nm^3}\right] - c_{SO2,load\ emit,refO2}\left[\frac{mg}{Nm^3}\right]\right)\cdot\frac{1}{f_{O2,corr}} \\ &\cdot \dot{v}^{flue\ gas}_{\lambda,dry,year}\left[\frac{Nm^3}{year}\right]\cdot 10^{-9}\ \left[\frac{tons}{mg}\right] \end{aligned} \tag{5-30}$$

Cost of lime consumption is provided by equation 5-31:

$$C^{LSD}_{CaO}\left[\frac{€}{year}\right] = \gamma^{s}_{CaO}\left[\frac{t_{\ CaO}}{t_{\ SO2}}\right]\ \cdot\ m^{LS\ D\ FGD}_{SO2\ abated}\ \left[\frac{t_{\ SO2}}{year}\right]\cdot\ C_{CaO,spe}\left[\frac{€}{t_{\ CaO}}\right] \tag{5-31}$$

CaO price is about 5 times price of limestone. Prices range from 80 to 250 €/t CaO according to its specific surface.

5.2.2.3. By-product, disposal or recovery

According to [Sargent 2007] by products can be used for aggregate and block manufacture or disposed. If not sold, by-products have to be disposed in correct conditions.

According to oxidation efficiency, $CaSO_3$ and/or $CaSO_4$, ½H_2O are produced. If the LSD FGD is installed in an existing plant, the existing ESP can be maintained and fly ashes can be eliminated. In new plants, the presence of deduster before the absorber is not systematic. The production of 50/50 $CaSO_3$ and $CaSO_4.½H_2O$ is assumed to calculate the by-product production (This explains the factor 2.366 in equation 5.32).

γ^{s}_{bp}: specific by-product production in ton by-product (bp)/ton SO_2 removed.

$$\gamma^{s}_{bp}\left[\frac{t_{bp}}{t_{SO2\ abated}}\right] = \gamma^{s}_{CaO}\left[\frac{t\ CaO}{t\ SO2}\right]\cdot\left(\frac{100-PUR_{CaO[\%]}}{100}\cdot 1+\frac{PUR_{CaO[\%]}}{100}\cdot 2.366\right)\left[\frac{t_{bp}}{t_{CaO}}\right] \tag{5-32}$$

Cost of waste disposal or gypsum sold is given by equation 5-33:

$$C^{LSD}_{bp}\left[\frac{\text{€}}{year}\right] = m^{LSD\ FGD}_{SO2\ abated}\left[\frac{t_{SO_2}}{year}\right]\cdot\gamma^{s}_{bp}\left[\frac{t_{bp}}{t_{SO_2}abated}\right]\cdot C_{bp,spe}\left[\frac{\text{€}}{t_{bp}}\right] \tag{5-33}$$

In case by-product is sold, $C_{bp,spe}$ is negative, indicating a benefit. In case of disposal, $C_{bp,spe}$ is positive. No information is available of the possible price of the product recovered if sold. When by-products are disposed, disposal prices depend on the waste disposal treatment. Landfill or other treatment such as incineration can be used. By-product disposal prices range from 0.33 to 89 €/t by-product, according to the information obtained.

5.2.3. Adaptation of costs for installations consuming heavy fuel oil

As carried out in the current EGTEI methodology, the investment function derived for coals is adapted for liquid fuels according to the relative flue gas volume, considering 1 for hard coal.

The average flue gas volume for coal at 6% O_2 is 358 Nm^3/MJ.

The average flue gas volume for heavy fuel oil at 3% O_2 is 285 Nm^3/MJ.

An average factor of 0.8 can be derived from those data. The investment functions presented above for coals are adapted with this factor, as follows.

The cost functions for the different modules of the LSD FGD for heavy fuel oil, are as follows:

- C^{LSD}_{abs} : absorber unit cost

$$C^{LSD}_{abs}[\text{€}\,] = 3{,}814{,}301\cdot r\cdot\left(cf\cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.4}\cdot\left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right]\right)^{0.01}\cdot(bs\ [MWe])^{0.716} \tag{5-34}$$

- $C^{LSD}_{reag,wh}$: reagent preparation and waste handling unit cost

$$C^{LSD}_{reag,wh}[\text{€}] = 389{,}586\ \cdot\ r\cdot\left(load_{SO2}\left[\frac{kg_{SO2}}{GJ}\right]\cdot GHR\left[\frac{GJ}{kWh}\right]\right)^{0.2}\cdot(bs[MWe])^{0.716} \tag{5-35}$$

- C_{bbp}^{LSD}: base balance of plan costs

$C_{bbp}^{LSD}[€\,] = 2{,}178{,}442 \cdot r \cdot \left(GHR\left[\frac{GJ}{kWh}\right]\right)^{0.4} \cdot (bs[MWe])^{0.716}$	5-36

The total equipment cost including retrofit factor $C_{equip}^{LSD}[€\,]$ is:

$C_{equip}^{LSD}[€\,] = C_{abs}^{LSD}[€\,] + C_{reag,wh}^{LSD}[€\,] + C_{bbp}^{LSD}[€]$	5-37

$C_{inst}^{ind}[€]$ includes engineering and construction management cost, labour cost for construction and contractor profit and fees and is estimated as a percentage of C_{equip}^{LSD} .

The total equipment cost including retrofit factor $C_{equip}^{LSD}[€\,]$ is provided by equation 5-6.

Other equations for total investment and operating costs do not change.

5.3. Dry FGD

The duct sorbent injection (DSI) FGD is considered for cost estimation. Dry hydrated lime ($Ca(OH)_2$), quick lime (CaO), sodium bicarbonate ($NaHCO_3$) and other sorbents can be used.

Often, due to high amount of un-reacted sorbent, recycling of products must be operated several times. The specific surface of sorbent is crucial for the efficiency of the process and sorbent prices depend on this characteristic. For sodium bicarbonate, recycling is not necessary due to its high reactivity. High efficiency may be obtained in a single pass process on a fabric filter [SOLVAIR 2014].

According to [EU commission, 2006a], the existing ESP can be kept in operation in case of retrofit (It is explained that the way an ESP is able to collect 100 times more product and fly ash than at a single throughput to a duct injection process, is not well understood). According to manufacturers of these technologies [SOLVAIR, 2014] using sodium bicarbonate, the use of a fabric filter is not mandatory. The ESP may be sufficient if it is able to comply with the ELVs when taking into account the additional dust load generated by the sodium sorbent injection and desulphurisation products. An efficiency of 80 % may be obtained on a single pass ESP.

The estimation of costs is carried out taking into account lime and sodium bicarbonate as reagents and assuming the presence of a new fabric filter after the existing ESP in case of retrofit. The FF mainly eliminates residual fly ashes and desulphurisation products. This is the technical solution proposed by most manufacturers of dry desulphurisation techniques [Coal WG, 2013] for existing plants. The residence time in the fabric filter can be enlarged to facilitate the desulphurisation. The air to cloth ratio of the FF has to be reduced consequently.

Dry FGD efficiency ranges from 50 to 80 % according to [EU Commission, 2006a]. Manufacturers of these technologies announce larger efficiencies up to 99 %, depending on sorbent quality (specific surface), arrangement made and excess of sorbent used.

5.3.1. Investment costs

The duct sorbent injection FGD is less documented than wet desulphurisation techniques. Information has been provided by experts from a French working group of the chemical industry [Coal WG, 2013] and manufacturers.

With assumptions made for the cost estimation, the main investment item is for the FF which has to be larger to enable finalisation of the desulphurisation. In case of retrofit, the existing ESP can be maintained. The FF eliminates desulphurisation products and a small fraction of remaining fly ashes when the ESP is present. From data provided by experts [Coal WG, 2013] and [SOLVAIR 2014], additional costs for sorbent injection in the ducts and sorbent preparation can be estimated to 30 % of the FF investment.

Sorbent demand depends on the deSOx efficiency required, the temperature, the load of SO_2 and the flue gas treatment layout (sorbent residence time before filter, duct design and quality of the solid-gas mix).

Hypothesis taken into account to derive the sorbent consumption according to the efficiency of the desulphurisation required, are as follows:

Lime:

At stoichiometry, 1 mole CaO abates 1 mole SO_2. In terms of mass, the molar ratio Ca/S = 1, accounts to CaO consumption of 0.875 t CaO/t SO_2 abated. [ADEME 2000] provides the following data:

Table 5.3: lime consumption factor for DSI FGD according to efficiency required for temperature between 130 to 140 °C

Efficiency of SO_2 removal η	t CaO/t SO_2 abated	Ratio Ca/S abated
30.0%	0.875	1
50.0%	1.750	2
70.0%	2.625	4

According to its characteristics, lime purity is about 95 to 97%. The purity has been introduced in the lime demand. A function has been derived to express the consumption according to the DSI FGD efficiency (equation 5-38).

The specific lime demand γ^{s}_{CaO}per t SO_2 abated is derived as follows, according to the DSI FGD efficiency required and lime purity:

$$\gamma^{s}_{CaO}\left[\frac{t\ CaO}{t\ SO2\ abated}\right]=\frac{\left(6.5625\ \cdot\ \frac{\eta^{dsi}[\%]}{100}-\ 1.2396\right)}{\frac{PUR_{CaO}\ [\%]}{100}} \tag{5-38}$$

γ^{s}_{CaCO3}: specific CaO demand in t CaO/t SO_2 removed

η^{LSD}: removal efficiency required for LSD

PUR_{CaO}: purity of the reagent CaO in %

According to oxidation efficiency and assuming humidification to enhance efficiency, $CaSO_3$ and/or $CaSO_4$, ½H_2O are produced. The production of 50/50 $CaSO_3$ and $CaSO_4$.½H_2O is assumed to calculate the by-product production (This explains the factor 2.366 in equation 5.39).

γ^{s}_{bp}: specific by-product production in ton by-product (bp)/ton SO_2 removed.

$$\gamma^{s}_{bp}\left[\frac{t_{bp}}{t_{SO2\ abated}}\right]=\ \gamma^{s}_{CaO}\left[\frac{t\ CaO}{t\ SO2}\right]\cdot\left(\frac{100-PUR_{CaO[\%]}}{100}\cdot 1+\frac{PUR_{CaO[\%]}}{100}\cdot 2.366\right)\left[\frac{t_{bp}}{t_{CaO}}\right] \tag{5-39}$$

Sodium bicarbonate:

At stoichiometry, 2 moles $NaHCO_3$ abate 1 mole SO_2. In terms of mass, this accounts for a $NaHCO_3$ consumption of 2.625 t $NaHCO_3$/t SO_2 abated. From data provided by [SOLVAIR 2014], the following data can be derived:

Table 5.3: Sodium bicarbonate consumption factor for DSI FGD according to efficiency required

Efficiency of SO_2 removal η	t $NaHCO_3$/t SO_2 abated	Ratio Na/S abated
70.0%	1.838	1.40
80.0%	2.192	1.67
90.0%	2.625	2.00
95.0%	3.150	2.40

According to its characteristics, technical grade sodium bicarbonate purity is about 98%. The purity has been introduced in the sorbent demand. A function has been derived to express the consumption according to the DSI FGD efficiency (equation 5-40).

The specific sodium bicarbonate demand γ^{s}_{NaHCO3} per t SO_2 abated is derived as follows, according to the DSI FGD efficiency required and the sorbent purity:

$$\gamma^{s}_{NaHCO3}\left[\frac{t\ NaHCO3}{t\ SO2\ abated}\right]=\frac{\left(4.9786\ \cdot\ \frac{\eta^{dsi}[\%]}{100}-\ 1.7185\right)}{\frac{PUR_{NaHCO3}\ [\%]}{100}} \tag{5-40}$$

$\gamma^{s}_{\mathrm{NaHCO3}}$: specific $NaHCO_3$ demand in t $NaHCO_3$/t SO_2 removed

η^{LSD}: removal efficiency required for LSD

PUR_{NaHCO3}: purity of the reagent $NaHCO_3$ in %

According to oxidation efficiency Na_2SO_3 is produced, then oxidised into Na_2SO_4 in contact with air.

γ^{s}_{bp}: specific by-product production in ton by-product (bp)/ton SO_2 removed.

$$\gamma^{s}_{bp}\left[\frac{t_{bp}}{t_{SO2\ abated}}\right]=\ \gamma^{s}_{NaHCO3}\left[\frac{t\ NaHCO3}{t\ SO2}\right]\cdot\left(\frac{100-PUR_{NaHCO3[\%]}}{100}\cdot 1+\frac{PUR_{NaHCO3[\%]}}{100}\cdot 1.690\right)\left[\frac{t_{bp}}{t_{NaHCO3}}\right] \tag{5-41}$$

The investment cost for the PJFF is estimated taking into account an air to cloth ratio, A/C, of 0.8 cm/s (to be in the range from 0.66 to 1 cm/s, see above), on average but may be increased up to

1,7cm/s with sodium bicarbonate. Investment equations are provided in chapter 4. The PJFF investment is increased by a factor 1.3 as explained above to account for equipment for sorbent injection and preparation.

The average by-product concentration to be abated in the PJFF is obtained by the following equation:

$$load_{bp,dry,in,\lambda}^{DSI\ FGD}\left[\frac{mg}{Nm3}\right]=\frac{m_{SO2\ abated}^{DSI\ FGD}\ \left[\frac{t_{\ SO2}}{year}\right]\cdot\gamma_{bp}^{s}\left[\frac{t\ bp}{t\ SO2\ abated}\right]}{\dot{v}_{\lambda,dry,year}^{flue\ gas}\ \left[\frac{Nm^{3}}{year}\right]}\cdot 10^{9}\ [\frac{mg}{t}] \tag{5-42}$$

$$load_{bp,dry,in,refO2}^{DSI\ FGD}\left[\frac{mg}{Nm3}\right]=load_{bp,dry,in,\lambda}^{DSI\ FGD}\left[\frac{mg}{Nm3}\right]\cdot\ f_{O2,corr} \tag{5-43}$$

If the ESP is not maintained, fly ash load has to be added (equation 2-1-14).

For the current development of the methodology, in the EXCEL tool, the situation with an ESP before the DSI FGD is assumed to estimate investment and operating costs.

5.3.2. Operating costs

Component of operating costs are as follows:

5.3.2.1. Electricity

Electricity consumption and costs are derived from equations presented in chapter 4.2.4.1. for the PJFF. Additional electricity consumption for lime injection has not been estimated due to lack of data.

5.3.2.2. Sorbent consumption

Lime demand is derived from equation 5-38, sodium bicarbonate demand from equation 5.40.

The amount of SO_2 abated is computed by equation 5-44:

$m_{SO2\ abated}^{DSI\ FGD} \left[\frac{t_{\ SO2}}{year}\right] = \left(c_{SO2,in,refO2}\left[\frac{mg}{Nm^{3}}\right] - c_{SO2,load\ emit,refO2}\left[\frac{mg}{Nm^{3}}\right]\right) \cdot \frac{1}{f_{O2,corr}} \cdot \dot{v}_{dry,year}^{flue\ gas}\left[\frac{Nm^3}{year}\right] \cdot 10^{-9}\ [\frac{t}{mg}]$	*5-44*

Cost of sorbent (lime or sodium bicarbonate) consumption is provided by equation 5-45:

$C_{Sorbent}^{DSI}\left[\frac{€}{year}\right] = \gamma_{Sorbent}^{s}\left[\frac{t_{\ Sorbent}}{t_{\ SO2\ abated}}\right] \cdot m_{SO2\ abated}^{DSI\ FGD}\left[\frac{t_{\ SO2}}{year}\right] \cdot C_{Sorbent,spe}\left[\frac{€}{t_{\ Sorbent}}\right]$	*5-45*

5.3.2.3. By-products, recovery or disposal

By-product production is derived from equation 5-39 for lime and 5-41 for sodium bicarbonate.

The total amount of by-product (or TSP) recovered by the PJFF is provided by equation 5-46.

$m_{bp,abated,PJFF}\left[\frac{t_{bp}}{year}\right] = \left(load_{bp,dry,O2ref}\left[\frac{mg}{Nm^{3}}\right] - c_{bp,load\ emit,refO2}\left[\frac{mg}{Nm^{3}}\right]\right) \cdot \frac{1}{f_{O2,corr}} \cdot \dot{v}_{\lambda,dry,year}^{flue\ gas}\left[\frac{Nm^3}{year}\right] . 10^{-9}$	5-46

Annual costs are defined according to equation 5-47:

$C_{bp}^{PJFF}\left[\frac{€}{year}\right] = \left(m_{bp,abated,PJFF}\left[\frac{t_{bp}}{year}\right] \cdot C_{bp,spe}\left[\frac{€}{t_{bp}}\right]\right)$	5-47

In case of by-product sold, C_{bp} is negative, indicating a benefit. In case of disposal, C_{bp} is positive.

5.3.2.4. Bag replacement

Bag replacement costs are derived from equation 4-12.

5.3.3. Fuel substitution

The mass of SO_2 abated by the use of low sulphur content fuel (compared to the fuel replaced) is calculated by equation 5-48:

$m_{SO2\ abated}^{low\ sulphur\ fuel}\left[\frac{t_{\ SO2}}{year}\right] = \left(load_{SO2,fuel\ 1}\left[\frac{kg_{SO2}}{GJ}\right] - load_{SO2,fuel\ 2}\left[\frac{kg_{SO2}}{GJ}\right]\right) \cdot cons_{e}^{fuel}\left[\frac{GJ}{h}\right] \cdot cap\frac{[\%]}{100} \cdot 8{,}760\ .\ 0.001\left[\frac{t}{kg}\right]$	*5-48*

Fuel 1 is the fuel replaced and fuel 2 is the new fuel with a lower S content.

Annual costs are due to additional costs of the low sulphur fuel:

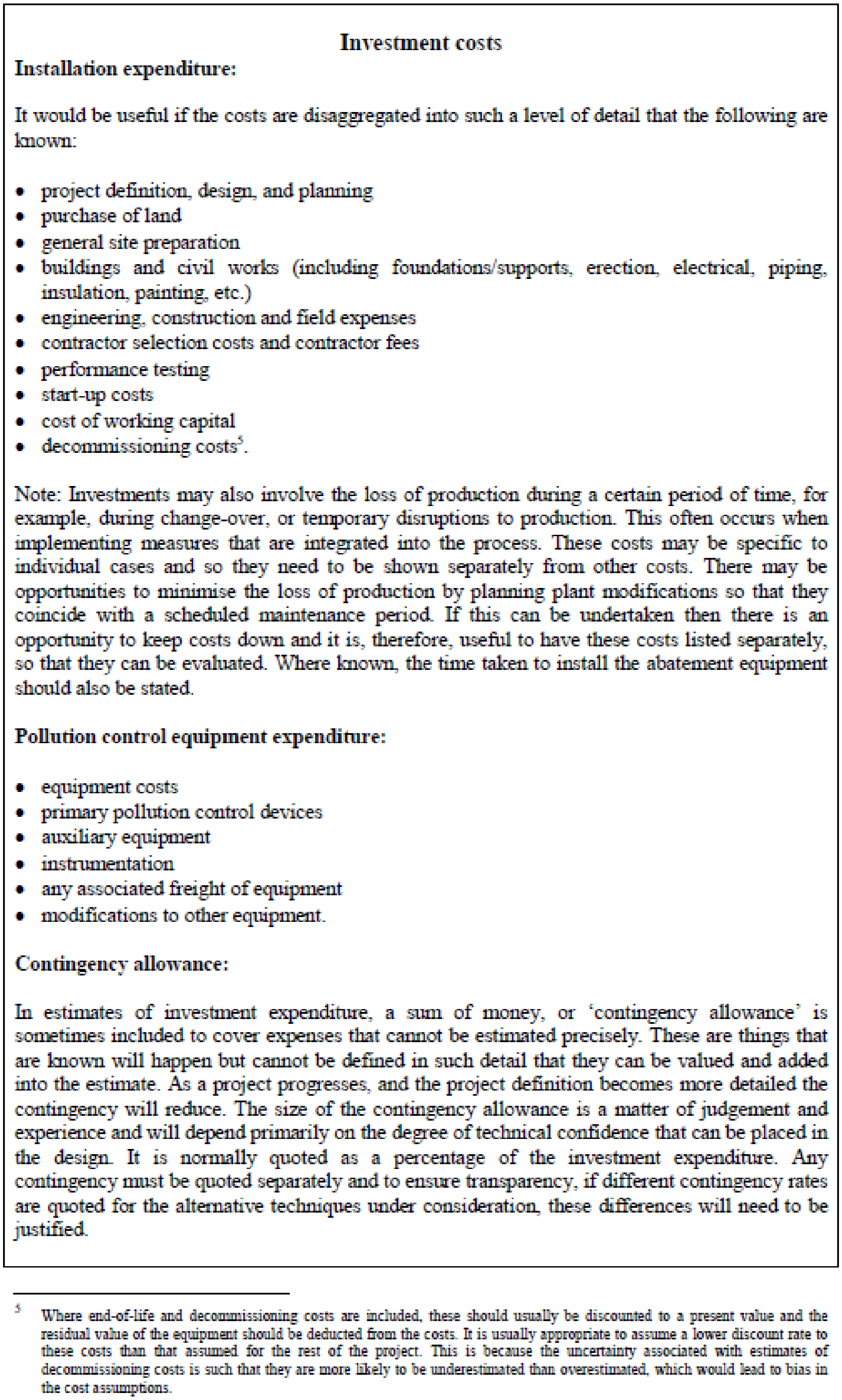
$$C^{low\ sulphur\ fuel}\left[\frac{€}{year}\right] = cons_e^{fuel}\left[\frac{t\ coal}{h}\right]\cdot\frac{cap\ [\%]}{100}\cdot 8{,}760\ \cdot\ C_{low\ S\ fuel,spe}^{additional}\left[\frac{€}{tcoal}\right] \tag{5-49}$$

Bibliography

ADEME 2000	Les techniques de désulfuration des procédés industriels
Balcke Dürr	Elektrofilter und Gewebefilter für kohlegefeuerte Kraftwerke im Vergleich, T.Riepe, 2008
Brandwood R.	EGTEI/UNECE abatement cost report. Joint environmental programme: JEP, 2008
Bunbury, H.M.	Die Chemie des Holzes, 1925
CCSD 2008	Analysis of pollution control cost in coal based electricity generation. Cooperative research centre for coal in sustainable development. 2008
Coal WG, 2013	Working group of the chemical industry for information exchange on reduction techniques for plants from 50 to 200 MWth, Chemical Industry Union, 2013
CR of 5 February 2013	Minutes of the EGTEI LCP cost update meeting of 5 February 2013
CUECost	Coal Utility Environmental cost (CUECost) workbook development documentation Version 5.0 ARCADIS for US EPA – EPA/600/5609/131, 2009
EDF 2013	Exchange with industrial experts
EGTEI 2013, 5/02/2013	Information delivered by Electricité de France (EDF) during the EGTEI “LCP Cost Group” meeting, 5th February 2013
EGTEI Questionnaires	Data collected from EGTEI questionnaires for 4 plants in Europe for collection of costs
E-On, Enel	Bag filters in power generation: drivers and considerations for the selection of a particulate control device 2009.
EU Commission, 2006a	Reference Document on Best Available Techniques for Large Combustion Plants, 2006
EU Commission, 2006b	Reference document on Economic and Cross Media Effects, EC, 2006
EU Commission, 2013	Draft 1 of the Reference Document on Best Available Techniques for Large Combustion Plants, July 2013
Evonik 2011	Evonik, 2011, ESP performance enhancements by SO_3 conditioning
Fabric Filter	http://158.110.32.35/CLASS/IMP-CHIM/Benitez-cap10.PDF ; available online 25-06-2013
Hamon, 2013	Information exchange meeting with Hamon, September 2013
Harding, N.S:	Biomass co-firing in utility boilers
IEA 2001	IEA coal research –Clean coal centre - Air pollution costs for coal fired power stations, 2001

Jedrusik, Swierczok, 2012	Jedrusik M., Swierczok A.: Design Efficiency of ESP, 2012. Wroclaw University of technology.
Jensen-Holm, H.; Thøgersen, J. R.; Lindenhoff, P.	Impact of biomass co-combustion on SCR DeNOx operation, 2009
Juda-Rezler; Kowalczyk	K. Juda-Rezler, D. Kowalczyk: Size distribution and trace elements contents of coal fly ash from pulverized boilers; Pol. J. Environ. Stud., 2012.
Koppejan, J., van Loo, S.	The Handbook of Biomass Combustion and Co-firing, 2010
Nalbandian, 2006	Economics of retrofit air pollution control technologies, 2006
Rubin	Modeling of integrated environmental control systems for coal-fired power plants. US department of Energy. 1991
Sargent 2007	Sargent & Lundy- Flue gas desulphurization technology evaluation Dry lime vs wet limestone FGD – March 2007
Sargent 2010a	Sargent & Lundy - SCR and baghouse capital cost estimate report, 2010. Salt river project
Sargent 2010b	Sargent & Lundy - Wet FGD cost development methodology – August 2010
Sargent 2010c	Sargent & Lundy - SDA FGD cost development methodology – August 2010
Smekens,K. (ECN)	Biomass co-firing in coal power plants in the Netherlands: effects on performance and air pollutant emissions,2013
SOLVAIR 2014	Information from the technical public documentation and specific information provided by SOLVAIR experts
Strauß 2006	Strauß, Karl: Kraftwerkstechnik, Berlin, Springer, 2006.
Termokimik corporation	The most performing existing filter on coal-fired power plants, Torrevaldaliga Nord (Italy): http://www.termokimik.it/pdf/torrevaldaliga_eng.pdf available online 15th October 2013.
US EPA 2002	Air Pollution Control Cost Manual, 2002
US EPA 2010	Documentation for EPA Base Case v.4.10 Using the Integrated Planning Model - chapter 5 http://www.epa.gov/airmarket/progsregs/epa-ipm/BaseCasev410.html
US EPA Lesson 3	US EPA, lesson 3 ESP design parameters and their effects on collection efficiency, 2.0-2/98
US EPA, lesson 4	US EPA Lesson 4 ESP design review. 2.0-2/98
Zevenhoven & Kilpinen	Particulates. Filter systems, 2001

Annex A



Investment costs

Installation expenditure:

It would be useful if the costs are disaggregated into such a level of detail that the following are known:

- project definition, design, and planning
- purchase of land
- general site preparation
- buildings and civil works (including foundations/supports, erection, electrical, piping, insulation, painting, etc.)
- engineering, construction and field expenses
- contractor selection costs and contractor fees
- performance testing
- start-up costs
- cost of working capital
- decommissioning costs[5].

Note: Investments may also involve the loss of production during a certain period of time, for example, during change-over, or temporary disruptions to production. This often occurs when implementing measures that are integrated into the process. These costs may be specific to individual cases and so they need to be shown separately from other costs. There may be opportunities to minimise the loss of production by planning plant modifications so that they coincide with a scheduled maintenance period. If this can be undertaken then there is an opportunity to keep costs down and it is, therefore, useful to have these costs listed separately, so that they can be evaluated. Where known, the time taken to install the abatement equipment should also be stated.

Pollution control equipment expenditure:

- equipment costs
- primary pollution control devices
- auxiliary equipment
- instrumentation
- any associated freight of equipment
- modifications to other equipment.

Contingency allowance:

In estimates of investment expenditure, a sum of money, or ‘contingency allowance’ is sometimes included to cover expenses that cannot be estimated precisely. These are things that are known will happen but cannot be defined in such detail that they can be valued and added into the estimate. As a project progresses, and the project definition becomes more detailed the contingency will reduce. The size of the contingency allowance is a matter of judgement and experience and will depend primarily on the degree of technical confidence that can be placed in the design. It is normally quoted as a percentage of the investment expenditure. Any contingency must be quoted separately and to ensure transparency, if different contingency rates are quoted for the alternative techniques under consideration, these differences will need to be justified.

---

[5] Where end-of-life and decommissioning costs are included, these should usually be discounted to a present value and the residual value of the equipment should be deducted from the costs. It is usually appropriate to assume a lower discount rate to these costs than that assumed for the rest of the project. This is because the uncertainty associated with estimates of decommissioning costs is such that they are more likely to be underestimated than overestimated, which would lead to bias in the cost assumptions.

Annex B – Exemplary investment data for DeNOx technologies

The following data of realised and estimated investments indicates the large range, a retrofit or a new investment into DeNOx technologies might have. When reading the data, it has to be borne in mind that most data reflects total project costs to the power plant operator and most projects have been retrofits. For the scope of retrofit (incl. necessary additional modifications), the type of costs which have been included and more details, please refer to the original literature.

		JEP	IEA	IEA	IEA	IEA
		LNB+SNCR	LNB	LNB	LNB	LNB
EUR / kWe	Min	28,51	2,35	5,83	5,83	17,14
	Max	28,51	2,35	20,83	30,00	17,14
EUR / kWth	**Min**	**71,29**	**5,88**	**14,58**	**14,58**	**42,85**
	Max	**71,29**	**5,88**	**52,08**	**75,00**	**42,85**
Given Literature Values	*Original Source*	*Barmor, 2006*	*Orfanoudakis, 2005*			*Schneck, 2003*
	Min	*20*	*5*	*7*	*7*	*15*
	Max	*20*	*5*	*25*	*36*	*15*
	Unit	*GBP/kWe*	*EUR/kWth*	*USD/kWe*	*USD/kWe*	*USD/kWe*
	Country	*UK*	*Greece*	*UK*	*US*	*Canada*
	Cost Year	*2007*	*2005*			*2003*
	Suppl. Info			*High End: OFA*	*High End: OFA*	

IEA	IEA	IEA	IEA	IEA	JEP	JEP	JEP	JEP
SCR	SCR	SCR	SCR	SCR	SCR	SCR	SCR	SCR
79,99	91,87	41,34	50,53	23,53	85,54	52,42	285,15	114,06
182,82	91,87	82,69	128,62	32,94	114,06	52,42	299,41	185,35
199,96	**229,68**	**103,36**	**126,33**	**58,82**	**213,86**	**131,04**	**712,87**	**285,15**
457,06	**229,68**	**206,72**	**321,56**	**82,35**	**285,15**	**131,04**	**748,52**	**463,37**
Morris, 2003				*Orfanoudakis, 2005*	*Barmor, 2006*	*Literature Survey*	*Literature Survey*	*Literature Survey*
70	*100*	*45*	*55*	*50*	*60*	*50*	*200*	*80*
160	*100*	*90*	*140*	*70*	*80*	*50*	*210*	*130*
USD/kWe	*USD/kWe*	*USD/kWe*	*USD/kWe*	*EUR/kWth*	*GBP/kWe*	*EUR/kWe*	*GBP/kWe*	*GBP/kWe*
US	*US*	*UK*	*US*	*Greece*	*UK*	*France*	*US*	*US*
2003				*2005*	*2007*	*2007*	*2007*	*2007*
Average 110 USD/kWe					*low eff. to high eff. SCR*		*Complicated*	

JEP	IEA		IEA	IEA
SNCR	SNCR		SCR	SCR
15,83	7,06		82,67	58,20
15,83	14,12		124,00	128,05
39,57	**17,65**		**206,66**	**145,51**
39,57	**35,29**		**310,00**	**320,12**
	Orfanoudakis, 2005		*US DOE, 2005*	*US EPA, 2002*
	15		*80*	*50*
	30		*120*	*110*
	EUR/kWth		*USD/kWe*	*USD/kWe*
	Greece		*US*	*US*
	2005		*2004*	*2001*
SNCR recalculated by subtracting the LNB average from LNB+SNCR Investment of Barmor, 2006				*US EPA Cost Methodology*

Annex C – Exemplary investment data for ESP and PJFF technologies

The following data of realised and estimated investments indicates the large range, a retrofit or a new investment into DeNOx technologies might have. When reading the data, it has to be borne in mind that most data reflects total project costs to the power plant operator and most projects have been retrofits. For the scope of retrofit (incl. necessary additional modifications), the type of costs which have been included and more details, please refer to the original literature.

Literature	**Cost Year**	**Currency year / power unit**					**EURO 2010 / kWth**	
		Min	Max	Currency		Power unit	**Min**	**Max**
RAINS	2005	17,30	17,30	£	/	kWe	11,90	11,90
AEP	2005	32,00	36,00	£	/	kWe	22,01	24,77
Nalbandian	2006	70,70	70,70	$	/	kWe	24,82	24,82
Naldandian [Orfanoudakis]	2006	85,00	85,00	$	/	kWe	29,84	29,84
World bank	2001	29,00	33	$	/	kWe	18,07	20,56
IEA	1998	40,00	60,00	$	/	kWe	20,18	30,28
Balcke Durr	2008	25,80	41,80	€	/	kWth	24,70	40,01
Questionnary Plant D	1983	13,30	13,30	€	/	kWth	23,33	23,33
Zevenhoven & Kilpinen	1982	9,65	9,65	$	/	kWth	17,28	17,28
Rubin	1985	43,00	43,00	$	/	kWe	38,32	38,32
Sankey	1997	19,00	19,00	$	/	kWth	23,86	23,86
IEA	1995	9,00	30,00	$	/	kWth	9,94	33,15

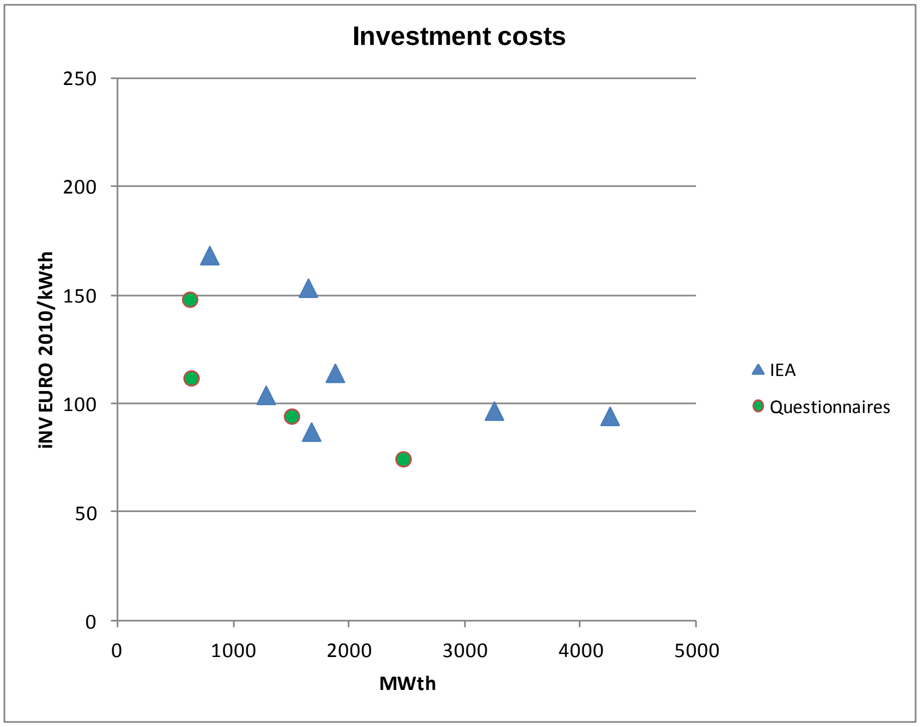
ESP investment costs

Literature	**Cost Year**	**Currency year / power unit**					**EURO 2010 / kWth**	
		Min	Max	Currency		Power unit	**Min**	**Max**
RAINS	2005	23,80	23,80	£	/	kWe	16,37	16,37
AEP	2005	28,00	28,00	£	/	kWe	19,26	19,26
Nalbandian [Smith]	2006	53,00	80,00	$	/	kWe	18,61	28,09
Naldandian [Orfanoudakis]	2006	60,50	60,50	$	/	kWe	21,24	21,24
Nalbandian	2006	60,00	60,00	$	/	kWe	21,07	21,07
World bank	2001	36,00	36	$	/	kWe	22,43	22,43
IEA	1998	50,00	70,00	$	/	kWe	25,23	35,32
Balcke Durr	2008	19,70	21,00	€	/	kWth	18,86	20,10
PM BART determination [American Electric Power]	2007	50,00	50,00	$	/	kWe	38,23	38,23
Sergent et Lundy	2010	65,21	87,56	$	/	kWe	18,74	25,16
IEA	2000	4,80	45,00	$	/	kWth	7,26	68,07
Wu	2001	3,90	16,50	$	/	kWth	6,08	25,72

PJFF investment costs

Annex D

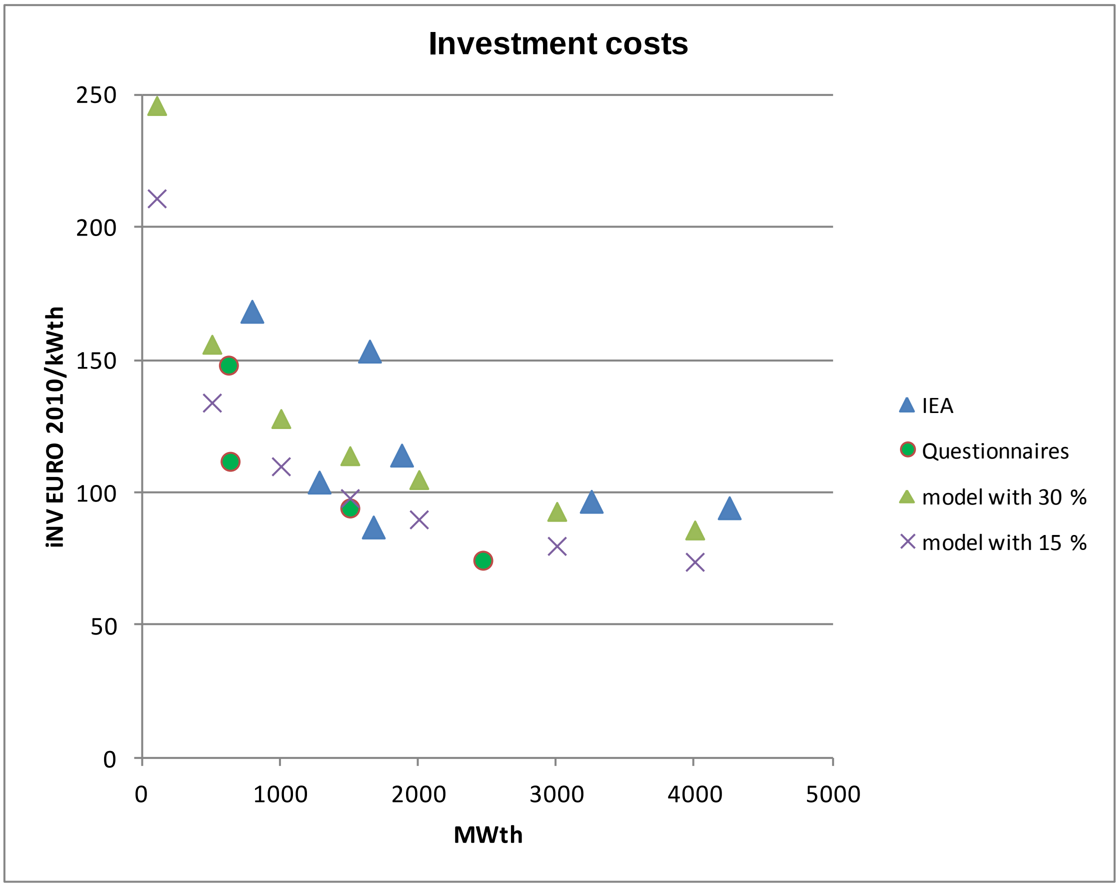
The following figure presents investment costs of wet FGD using limestone and producing gypsum (IEA 2001, questionnaires).



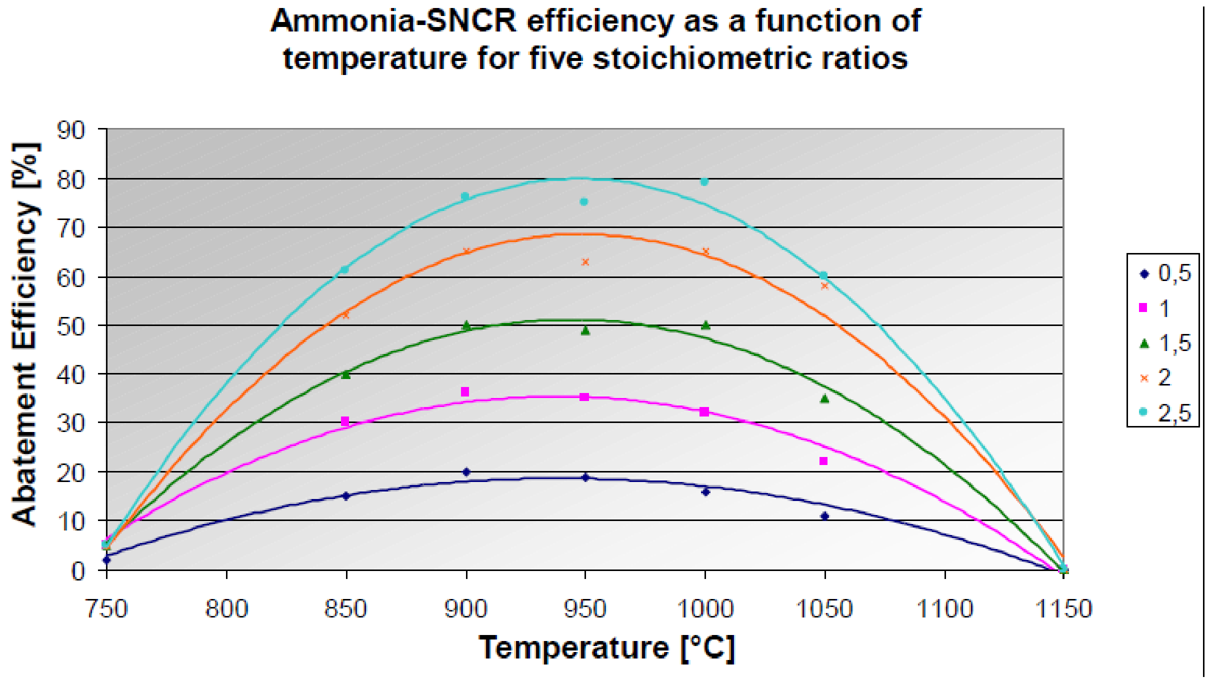
The cloud of points demonstrates that investments are very site specific. Various parameters can influence costs. Investments presented above are for existing plants. The difficulty of the retrofit is not known. The sulphur content of coals can be different, the year of construction can be different (all costs are expressed in Euro 2010). Other technical parameters of the wet FGD may also influence the investment.

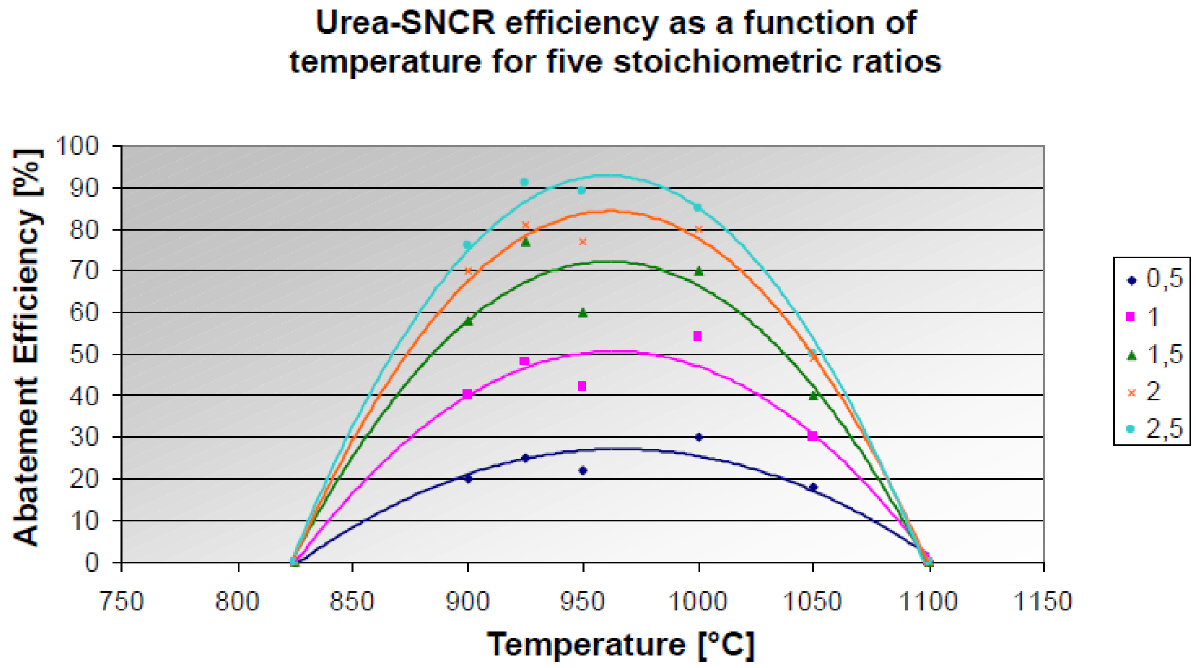
The bibliography study carried out by the EGTEI technical secretariat and discussions during the previous technical meetings of the EGTEI sub-group on LCP have concluded that the cost model developed by the US EPA based on works of Sargent and Lundy [Sargent 2010b] could be used.

The investments estimated according to chapter 5.1 and equations 5.1 to 5.8 are presented in the following figure. The value of one parameter used in equation 5-7 (15 or 30 %) can influence the total investment. The ratio 30% has been selected as it seems to be in the range of all costs collected.



Annex E – Interdependence of SNCR operating parameters and the abatement efficiency


Ammonia-SNCR efficiency as a function of temperature for five stoichiometric ratios


Urea-SNCR efficiency as a function of temperature for five stoichiometric ratios